Chemical Engineering Progress

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Opinion and Comment

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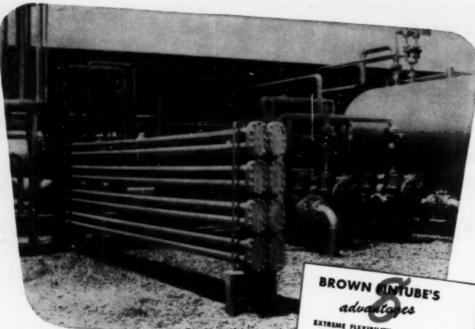
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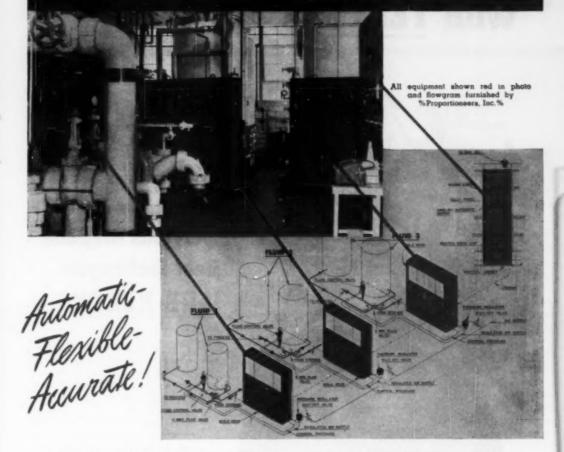
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OPINION AND COMMENT

"SOMETHING FOR NOTHING"

IN a letter to his friend Randall, on May 23, 1857, Lord Macaulay predicted that our government could never survive the Twentieth Century. His reasoning was that the indolent and unthinking people of our nation would be in the majority and would, sooner or later, elect government representatives who would take from the rich and give to those who elected them. Taxation would be increased and liberty would be retracted. In the end, either a modern Napoleon would seize control of our government or our nation would be plundered and laid waste by barbarians developed within our own country by our own institutions.

Such a prediction is disturbing when considered in the light of the rapidly spreading philosophy of "something for nothing." People desire security but also the fruits derived from risks that others have taken at the expense of security; people want higher wages, yet they want to do less work; they want a higher standard of living, yet they want to produce less; the farmers demand a higher than market price for their produce, which entails subsidizing them. Yes, the majority seems to believe, at least subconsciously, that it is possible to get more for less or something for nothing. They do not seem to realize that we enjoy the highest standard of living in the world because our production rate per capita of useful products is the highest in the world—not because we have legislated it or devalued the dollar. Nor do they understand that our security, both individually and collectively, is dependent upon our ability to produce something useful rather than upon any legislation or manipulation of our money system.

Anything which curtails the production rate of useful products tends to reduce our standard of living and to affect our security adversely. We are familiar with how the Egyptian economy was destroyed by building pyramids. Of course we are not building pyramids, but some things we are doing are less useful and if continued indefinitely will have the same effect.
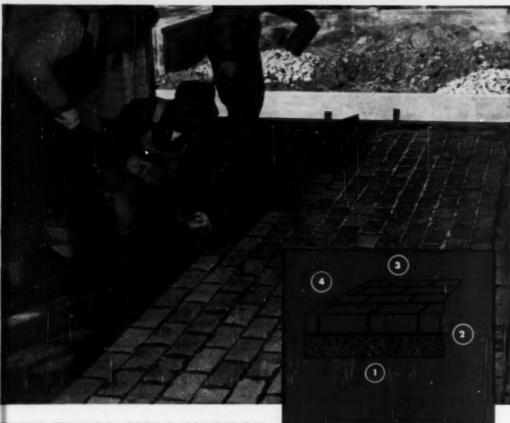
Perhaps it is wishful thinking but I cannot believe that the American people will continue, to the bitter end, to follow others up the blind alley in the construction of modern pyramids in the pursuit of something for nothing. Surely the American people will soon exercise their good judgment and common sense by weighing this philosophy on the balance of sound principles. For only a relatively short period can the majority reap benefits by actions which violate sound principles. Eventually all the seed corn will have been eaten and our children, if not we, will pay for our blunders.

It is my opinion that the people will eventually "get wise" to themselves that strikes are wasteful and a better method to settle labor disputes will be established. For obvious reasons, labor leaders and professional politicians, as well as some industrialists have objected to a labor court. I feel that the people will eventually come to the conclusion that we cannot legislate indefinitely security or prosperity. I believe that the people will finally learn that they can spend their money more intelligently than can the professional politicians. They will then demand a large reduction in taxation.

Every red-blooded American should always resist any increase in per capita taxation and any encroachment upon personal liberties regardless of how small they may be. Socialism can be established by means of increasing per capita taxation which, of course, entails retrenchment in personal liberty in favor of the professional politicians. If it were put to a vote and you would vote against establishing a socialistic government, then you had better give our tax situation some concentrated thought. We are headed in the direction of socialism by default.

Chemical engineers need not be convinced that there is no such thing as something for nothing. We have been well grounded on basic principles. We know that something cannot be created out of nothing. Even so, I am amazed at times to hear chemical engineers ask, "What will I get out of a membership in A.I.Ch.E.?" They should be asking, "What means does the Institute provide for me to contribute to my profession?" Everything we get out of our profession was contributed by someone who was not looking for something for nothing. Chemical engineers have a responsibility to their country to correct the perverted thinking of those fellow-citizens who, at least subconsciously, are looking for something for nothing.

C. G. Kirkbride

ENGINEERING AND ECONOMIC EVALUATION OF PROJECTS

. . . chemical engineers must begin their evaluation of projects long before they reach the pilot plant stage . . . an important step in project evaluation comes after laboratory work has been initiated or completed and before pilot plant work begins . . .

E. C. DYBDAL

Monsanto Chemical Company

Dayton, Ohio

CHEMICAL engineers and chemists are familiar with the general course of a new chemical development from the conception of the idea to the final manufacture of the product. After an investigation in the laboratory under the domain of the research chemist, the project emerges and becomes largely the responsibility of chemical engineers who guide it through various stages of process development and pilot plant investigation into plant production.

Although traditionally the sphere of interest of chemists is the research laboratory and that of chemical engineers is process development and plant production, research projects that reach the production stage are invariably the joint effort of chemists and chemical engineers in their evolution. After the laboratory work is completed, most chemists remain in close touch with their projects until all process development and pilot work is finished and plant operation has been proved. Likewise, the chemical engineers' contact with a project may begin long before process development and pilot plant work is started. Questions involving engineering practice and economics constantly arise. To answer these questions,

chemical engineers are called upon to evaluate projects at four more or less distinct stages in their development.

The first stage for many projects is an engineering and economic analysis of available data before any laboratory work is attempted. The object of this analysis is to determine the potentialities of a project for further research and development work, and for eventual exploitation.

The second stage is reached after some or all of the initially planned laboratory work has been completed. At this point enough information is available for a satisfactory material balance and for a preliminary analysis of the unit operations and processes involved, although several assumptions may be necessary to make up for the data that are lacking. The evaluation completed at this stage should provide the basis for a decision in regard to the advisability of going ahead with additional laboratory work or with process development and pilot plant work.

At the third stage the program of laboratory, bench scale, and pilot plant work has been completed. Where necessary, the material balances have been proved with larger-sized batches. Unit

operations and processes have been thoroughly studied and, if possible, all information needed for the design of the manufacturing plant has been obtained. A firm cost estimate and economic evaluation then are made in order to learn whether the project is still economically sound before the expense of the final engineering design is incurred.

The fourth stage comes after completion of the detailed plant design. At this time the cost estimates made previously are checked with values derived from the final design before funds for the manufacturing plant are appropriated.

Why Evaluate Projects?

Although the same general principles of engineering and economic analysis apply to all four of these stages, this discussion will be confined to the second or research stage. What is the purpose of estimates based on research laboratory investigations? How can preliminary evaluations and estimates contribute to the efficiency of research laboratory operations? How do they help to shorten the over-all time required to move a project from the research lab-

oratory into commercial production? How are they made and what are the problems encountered? These are the questions to be considered.

The principal reason for preparing engineering and economic evaluations of projects at the research stage is to detect, as early as possible, projects that are economically unsound or, impractical. In case two or more processes for the manufacture of a chemical are feasible, evaluation must show which of the processes is preferred so that the research work can be concentrated on this process.

Another object is to learn, at an early date, the engineering and design problems involved. Through circulation of the written report on the evaluation, other chemical engineers and chemists may become aware of these problems and the combined knowledge of several individuals may be focused on the project.

If the engineering and economic evaluation shows that the project is an interesting one, important additional information can be developed. Major items in the manufacturing cost are ascertained and methods for reducing them suggested. Additional information that will be required and its source, i.e., in laboratory, bench scale or pilot plant operations may be noted. An approximate minimum cost for obtaining the required laboratory, bench scale, or pilot plant work can be estimated.

The importance of evaluating laboratory projects as early as possible should be stressed further. The engineering and economic analysis should begin as soon as a reasonably reliable material balance is available and the rough details of the processing and purification steps are outlined. Frequently, research reports are handed to the chemical engineer after a detailed program of laboratory work has been completed and the laboratory phase of the work is finished. When this happens he finds, in some cases, that the project is uneconomical and that this fact could have been recognized earlier. Considerable effort might have been saved or directed into other channels. If the project proves to be attractive he may find that some additional information needed could have been obtained most advantageously before the laboratory work was halted. The cooperative effort of the chemical engineer and the chemist is needed to get the most information out of the laboratory.

From the viewpoint of the research chemist, early evaluation of projects at the research stage is of considerable importance. The chemist is interested in knowing whether he is wasting his time on a project that can be shown to be of little interest to the company no matter how ingenious his approach may be. By pursuing an active program of engineering and economic evaluation along with the laboratory work, research chemists will be able to handle more projects and to devote more of their effort to purely chemical investigations.

Analysis of Laboratory Projects from Engineering Standpoint

An engineering evaluation of projects at the research stage is designed to achieve the following:

1. Furnish information for an economic analysis of project.

2. Outline unit operations and chemical processes involved in projected manufacturing operations.

3. Show what additional information will be needed to complete design of a plant and how obtainable.

To accomplish these objectives the chemical engineer must first obtain basic laboratory data from the research chemist. A form such as the one shown in Figure 1 may be used to aid in assembling these data. Use of a form has an additional advantage; the chemist has before him a tabulation of the kind of data the chemical engineer will require and he can fill in the information during his investigation.

Not all details specified in Figure 1 need to be available before a preliminary evaluation is started. It may be advisable to begin evaluation as soon as a proved material balance and a brief outline of the reaction and purification steps are ready. At this point the yields may not be too well defined. The actual yield obtained is used but it may also be desirable to calculate raw material requirements on the basis of the highest yield visualized or on a 100% yield to show the effect of an increased yield on the cost. Many assumptions will be required at this early stage but these can be checked as the laboratory work progresses if the subsequent economic evaluation shows that it is worth while to do so.

After the laboratory data have been assembled the chemical engineer may sketch first a detailed flow sheet showing the main items and such auxiliary equipment as pumps, conveyors, scales, storage tanks, and bins. It is good practice to allow for these auxiliary items in preliminary evaluations although they may account for only a small part of the equipment cost.

The next step will be to obtain from management the number of pounds a year that may be required and the location of the plant. The chemical engineer can then size the various items of equipment and calculate the quantities of steam, electricity, fuel gas or oil, cooling and process water, etc., that will be needed, noting at each step the assumptions made. In sizing and specifying equipment for the various unit operations and processes, the chemical engineer will usually have to consult frequently with others in the company, and use the chemical engineering literature to obtain the best plant layout.

A written description of the projected manufacturing operations, unit operations and chemical processes involved, the assumptions, data that will be required for process design and how they may be obtained, completes the engineering evaluation. This information will become part of the report covering both the engineering and economic evaluation of the project.

Analysis of Laboratory Projects from Economic Standpoint

The foregoing investigation of the engineering aspects of a project serves as the basis for the economic evaluation which, briefly, is designed to answer these questions:

1. How much will it cost to manufacture the product at the desired rate?

2. How much capital will be needed for the new venture?

3. What must the selling price be to give a satisfactory return on total capital investment?

The general procedure for preparing economic evaluations consists in estimating the following:

1. *Total manufacturing cost.* The material balance, flow sheet, equipment sizes, and utility requirements developed for the engineering evaluation serve as the starting point (Table 1).

2. *Total capital investment.* This includes capital for the manufacturing machinery and equipment, nonmanufacturing M & E, and working capital (Table 4).

3. *Estimation of a selling price.* This gives a definite return on total capital investment after accounting for all costs involved, or conversely, the return that can be realized from selling product at a fixed price (Table 5).

Details for these procedures will be discussed in the following sections.

Fig. 1.

BASIC LABORATORY OR BENCH SCALE DATA

FOR THE PRODUCTION OF _____

PRINCIPAL REACTIONS: _____

SECONDARY REACTIONS: _____

PHYSICAL AND CHEMICAL PROPERTIES OF —

COMPOUND	_____ _____ _____
FORMULA	_____ _____ _____
MOLECULAR WT.	_____ _____ _____
BOILING POINT	_____ @ _____ _____ _____
VAPOR PRESSURE	_____ @ _____ _____ _____

DATA ON REACTION STEP OR STEPS FROM A TYPICAL RUN

MATERIAL BALANCE

RUN NO. _____

MATERIALS IN:

_____STEP			_____STEP			_____STEP		
COMPONENT	gms.	gm. MOLES	COMPONENT	gms.	gm. MOLES	COMPONENT	gms.	gm. MOLES

MATERIALS OUT:

REACTION CONDITIONS:
   TEMPERATURE —
   PRESSURE —
   RETENTION TIME —

% CONVERSION:

$$\frac{\text{gm. MOLES PRODUCT}}{\text{gm. MOLES REACTANT CHARGED}} \times 100 =$$

% YIELD:

$$\frac{\text{gm. MOLES PRODUCT}}{\text{gm. MOLES REACTANT USED UP}} \times 100 =$$

MATERIALS IN (OR OUT) BY PHASES FOR THE _____ STEP

MATERIALS _____(IN OR OUT)			MATERIALS _____(IN OR OUT)			MATERIALS _____(IN OR OUT)		
_____PHASE			_____PHASE			_____PHASE		
COMPONENT	gms.	gm. MOLES	COMPONENT	gms.	gm. MOLES	COMPONENT	gms.	gm. MOLES

PRINCIPAL PURIFICATION STEPS (WASHING, DISTILLATION, CRYSTALLIZATION, FILTRATION, ETC.)

MATERIALS IN:

_____STEP			_____STEP			_____STEP		
COMPONENT	gms.	gm. MOLES	COMPONENT	gms.	gm. MOLES	COMPONENT	gms.	gm. MOLES

MATERIALS OUT:

CONDITIONS FOR _____ PURIFICATION STEP
   TEMPERATURE _____
   PRESSURE _____
   TIME _____

BULK DENSITIES: (REACTANTS, FILTER CAKES, PRODUCT, ETC.)

OVERALL CONVERSION: (REACTION AND PURIFICATION STEPS)

$$\frac{\text{gm. MOLES PRODUCT}}{\text{gm. MOLES REACTANT CHARGED}} \times 100 =$$

OVERALL YIELD: (REACTION AND PURIFICATION STEPS)

$$\frac{\text{gm. MOLES PRODUCT}}{\text{gm. MOLES REACTANT USED UP}} \times 100 =$$

DESCRIPTION OF PROCESS, BLOCK AND LINE FLOW SHEET, AND ADDITIONAL NOTES —

(CONTINUE ON OTHER SIDE OF SHEET.)

_____ CHEMIST

Preparation of Preliminary Production Cost Estimates

Basic principles of economic analysis applied to the chemical industry are fully developed by Tyler (*11*). A general exposition of accounting procedures applied to chemical industry has been presented by Sheehan (*10*).

The first step in an economic analysis is estimation of the manufacturing cost of the product. The usual form for manufacturing cost estimates is shown in Table 1. When this table is used for preliminary estimates, many items are calculated by means of simple ratios which give sufficient accuracy in view of the uncertainties involved. However, for the firm cost estimates made after the process development and design are completed (third and fourth stages of development described earlier), formulas supplied by the accounting department are used for estimating many items in Table 1. Values from a typical preliminary cost estimate are shown in Table 1 as examples in the discussion of the estimating procedures. Taken consecutively they are:

Production Rate-operating Time. A plant running full time would be on stream for 8,760 hr./yr. Since plant equipment has to be shut down periodically for repairs, catalyst changes, holidays, etc., a conservative operating time for many processes is 8,000 hr./yr.

Capital—Machinery and Equipment. Capital—M & E. means the capital needed for the production machinery and equipment, installed and ready for operation. Capital required for plant machinery and equipment is estimated according to the general form shown in Table 2. The first requirement is information on the price of the individual items of equipment which are numbered on the flow sheet. This may be obtained, in most cases, from the company engineering files or may be estimated from published data (*2-3, 8, 10-11, 13-14*).

After the delivered price of the equipment is decided, cost of erecting and setting the equipment in place is estimated. This usually runs from 5 to 30% of the delivered price for most of the items; 15% is a fair average. The sum of the delivered cost of the equipment and the labor of erection is shown as the total numbered equipment cost in Table 2.

In order to obtain the total installed cost of the machinery and equipment, it is necessary to estimate the cost of piping, wiring, instruments, etc., listed in Table 2. For preliminary estimates, these items are calculated usually as percentages of either the total plant cost

---

TABLE 1.—MANUFACTURING COST ESTIMATE

Product .... X .... in .... (drums, bags, tank cars, etc.)

Production Rate—5,000,000 lb./yr. of 8,000 hr., 333 days

Capital—M & E $392,000

—Building, $ 50,000

Plant Location ....

Raw Materials	Unit	Quantity a Year	Unit Cost ($)	$ a Year	$ a cwt.
Sub-total					
Credits					
Net R.M.				792,460	15.85

Direct Conversion Expense

Labor ....man-hour		23,150	1.48	34,260	0.69
Supervision				6,000	0.12
Payroll charges				7,250	0.15
Steam ....M lb.		32,900	0.86	27,500	0.55
Electricity ....kw.hr.		1,701,900	0.0104	17,700	0.35
Fuel ....therms		63,600	0.0275	1,750	0.04
Compressed air ......M cu.ft.					
Refrigeration ....tons					
Inert gas ....M cu.ft.					
Water-cooling ......M gal.		409,100	0.022	9,000	0.18
Water-process ......M gal.		8,000	0.0802	640	0.01
Repairs				33,200	0.66
Factory supplies				1,920	0.04
Clothing, laundry				280	0.01
Laboratory				3,000	0.06
Royalty					
Total				142,480	2.86

Indirect Conversion Expense

Depreciation, M & E				39,200	0.78
Depreciation, building				1,500	0.03
Taxes				8,680	0.17
Insurance				1,060	0.02
Controllable indirect conversion expense				28,500	0.57
Total				78,940	1.57
Bulk conversion cost					4.43
Bulk manufacturing cost					20.28

Packaging and Shipping

Containers				26,500	0.53
Packing labor				3,200	0.07
Shipping supplies and labor				3,800	0.08
Freight allowances					
Total				33,600	0.68
Total manufacturing cost, f.o.b.				1,047,460	20.96

**TABLE 2.—MANUFACTURING MACHINERY AND EQUIPMENT ESTIMATE
NUMBERED EQUIPMENT LIST**

Item No.	Number Required	Description	Material	Labor	Total
Total					
			$108,800	$17,700	$126,500

Total numbered equipment	$126,500
Piping	61,700
Wiring	25,000
Instruments	20,000
Insulation	5,000
Foundations, supports and platforms	20,000
Site preparation, painting, and cleanup	8,000
Miscellaneous	4,800
Sub-total	281,000
Project overhead and contingencies	111,000
Total manufacturing M & E	$392,000
Manufacturing buildings	$ 50,000

$$\frac{\text{Total manufacturing M \& E}}{\text{Total numbered equipment}} = 3.1$$

or of the delivered equipment cost. Published information on these percentage factors has been summarized (8).

A short method for obtaining the total installed cost of the machinery and equipment consists in taking the sum of the delivered cost of these items, adding the cost of erecting and setting them in place (total numbered equipment in Table 2), and multiplying by a factor. A somewhat different factor is used to get the total cost of the plant including the manufacturing buildings. Development of factors for this purpose is discussed by Lang (7), Harris (4) and Nelson (8). (Note in Table 2 that the ratio of the total M & E to the total numbered equipment is 3:1.) For the average chemical process where operations with solids and fluids are conducted, the total M & E will usually be about three times the total numbered equipment. The estimator may vary the factor, possibly along the lines suggested by Lang (7), for processes dealing with solids, solids-fluids, or fluids. Use of factors for obtaining the total installed M & E capital is sufficiently accurate for many preliminary estimates. Greater accuracy will result from their use than could be obtained by an inexperienced estimator. (See Footnote.*)

Capital—Building. Cost of buildings is estimated on the basis of the cubic feet of volume or of the number of square feet of floor area (14). With present-day inflated building costs, a combination steel and masonry building for chemical processing may range from $7.00 to $12.00/sq.ft. or $0.20 to $0.30/cu.ft. These prices are for the shell of the building only; the building with services such as water, electricity,

* Since this paper was presented, a valuable correlation of process equipment cost data has been published by Chilton (2). The installed cost of many items of process equipment, adjusted to ENR index = 400, is plotted against size or capacity. Installed cost includes basic cost plus installation labor, foundations or supports, installation of auxiliary equipment, insulation, painting, and any piping considered an integral part of the equipment. An estimate of the reliability of the cost data is given in each case.

From the installed cost of equipment, total plant cost may be obtained as follows: (1) By adding to the installed cost the sum of percentage factors for process piping, instrumentation, outside lines, manufacturing buildings, and auxiliary facilities to obtain total physical cost. (2) By applying factors for engineering and construction costs, contingencies, and size to total physical cost to obtain total plant cost.

Total plant cost is then adjusted by multiplying the ratio of the prevailing ENR index to the base ENR index of 400 to obtain the current plant cost.

sprinklers, etc., installed, may run 50% higher. Approximate costs for various kinds of buildings have been summarized by Nelson (8).

Plant Location. Since cost of utilities: steam, power, and cooling water, varies throughout the country, location of the projected plant should be stated on the manufacturing cost estimate sheet.

Raw Materials. Fairly representative prices of most commercially available chemicals can be found in *Oil Paint and Drug Reporter* or in *Chemical and Engineering News*. For firm production cost estimates, the prices listed in these journals should be verified by consulting the manufacturers to make sure that they apply to the quality and quantity of the individual chemicals required in the projected manufacturing operations. Prices for chemicals not listed in the journals must be obtained from the manufacturers.

A chemical engineer may obtain, from sources within his company, current market prices of chemicals the company purchases for its manufacturing operations as well as interdepartmental or divisional transfer prices of chemicals produced and consumed within the company. Such price information, of course, is not available to individuals outside the company.

Credit for by-products varies with each particular case. Generally, by-products are credited at their sales price minus additional selling, shipping and purification costs, if any. By-product steam and power are credited to show the magnitude of these items compared with steam and power requirements shown under the direct conversion expense.

Direct Conversion Expense

Labor. Labor rates at the proposed location of the plant are usually available. Happel, Aries, and Borns (3) give rates for various grades of labor in effect December, 1946. These rates are as follows:

Leaders and foreman ..$1.65-$2.50/hr.

Skilled labor ....$1.25-$1.65/hr.

Unskilled labor ....$0.85-$1.25/hr.

If approximate wage rates are desired, these values may be increased by about 20% to apply to present-day conditions. The unit cost of labor shown in Table 1 is an average for the various classes of labor involved.

To allow for time lost in shutdowns, the labor expense should be estimated on the basis of 8,500 hr./yr. if the scheduled operation calls for 8,000 hr.

Supervision. An estimate is made of the number of hours of supervision required in normal manufacturing operations. The extra supervision needed during initial runs with the equipment is generally charged to starting-up expenses. A supervisor's salary may range from $5,000 to $6,500/year.

Payroll Charges. Payroll charges cover such items as social security, vacations, pensions, and compensation insurance. They may be estimated as a percentage of labor plus supervision. The usual range is from 12% to 20% of the sum of these items.

Steam. Steam requirements are determined by taking the total needed for the items of equipment in the manufacturing plant and adding the estimated amount required for building heat. Vilbrandt (14) shows how heat required for buildings may be estimated. For a rough estimate it is satisfactory to add about 25% of the steam required in the manufacturing operations to allow for radiation, line losses and building heat. The sum of the process steam and building heat is multiplied by a factor ranging from 1.25 to 2.0, depending on the steam load and nature of the operations, in order to take care of contingencies and losses.

Steam costs range from $0.25 to $0.90/M lb.

Electricity. Power requirements for process equipment and motors are calculated allowing for line losses. The lighting load is estimated. The sum of these two should be multiplied by a factor of 1.10 to 1.25 to allow for contingencies. Industrial power rates range from $0.0035 for locations having cheap power to $0.014/kw.hr. where generation and distribution costs are higher.

Fuel. Fuel gas ranges from $0.01 to $0.06/therm. An approximate value for fuel oil on a therm basis is $0.06.

Compressed Air. Cost of compressed air ranges from $0.03 to $0.10/M cu.ft. depending on pressure and size of the compressor required.

Refrigeration. Refrigeration costs vary widely and depend on a number of factors, chiefly, power, cooling water, depreciation and repairs. Because of the

interdependence of these factors and their variation from one installation to another it is necessary to estimate refrigeration costs for each installation if a reliable cost is required.

The following approximate refrigeration costs for single-stage ammonia compression systems are given to show, in a general way, how refrigeration costs increase with decreasing evaporator temperatures for installations of roughly 10 to 500 tons in size. In developing these costs, power was taken at $0.01/kw.hr., and 85° F. cooling water at $0.02/M gal.

Evaporator Temp. ° F.	$ a ton-day (or 288,000 B.t.u. abstracted)
34	0.47
0	0.82
—17	1.07

Steam jet refrigeration is widely used to obtain 50° F. cooling water. If steam is taken at $0.25/M lb. and 85° F. cooling water at $0.02/M gal., cost of steam jet refrigeration will be approximately $0.52 a ton-day, or 288,000 B.t.u. abstracted under normal operation. Steam jet refrigeration is especially advantageous where cheap low-pressure exhaust steam and cooling water are available.

Although temperatures below —30° F. are becoming increasingly common, sufficient data have not been published to develop approximate costs.

In preliminary estimates it is convenient to include the cost of refrigeration units in the manufacturing M & E and to charge the power, cooling water, repairs, and depreciation pertaining to refrigeration units, under these items in the conversion cost, if refrigeration costs for the particular plant location are not available.

Inert Gas. Nitrogen and carbon dioxide, in various degrees of purity, are most frequently used to provide inert atmospheres. Commercially pure nitrogen may be purchased, delivered in trailers, for the following approximate cost/100 cu.ft.

(1 atm. and 70° F.) consumed in one month:

cu.ft./month	$/100 cu.ft.
0- 80,000	0.65
80,000-150,000	0.60
150,000-	0.55

The above costs will vary depending upon delivery charges.

Cost of using liquid carbon dioxide has been estimated to be $3.00/100 lb. compared with $1.00/100 lb. for inert gas generated at a rate of 3,000 cu. ft./hr. by burning natural gas with air (1).

Water-cooling. The total quantity of cooling water required a year is calculated and multiplied by a factor of 1.10 to 1.50 to allow for losses and contingencies.

Cooling water costs vary greatly depending on the geographical location, water-treating problems, pumping costs, etc. If it is necessary to use well water the cost may range from $0.02 to $0.25/M gal. River and salt water costs usually range from $0.01 to $0.02/M gal. Reclaimed cooling tower water costs approximately $0.01 to $0.03/M gal.

Water-process. Process water is used in chemical reactions and in washing, extracting, dissolving, and similar processing operations. It also includes water used for drinking, for sanitary facilities, and for general clean-up and washing. Estimates should be generous for the latter requirements.

Fresh water, treated or untreated well or city water, is used. In some cases distilled or deionized water may be required. The cost of process water is generally with the range of $0.02-$0.25/M gal. Cost of deionizing water is compared with the cost of distilling by Kominek (6).

Repairs. For preliminary estimates repairs can be taken as a percentage of the installed plant machinery and equipment cost. Repair charges range from 5% to 15% of the machinery and equipment cost. A consideration of the type of equipment and operations involved will show whether repair charges should be high, low, or average. Repairs on buildings may be about 5% of their cost. Methods of handling repair charges are reviewed and a new method is proposed by Pierce (9).

Factory Supplies. This category includes items such as lubricants, janitor supplies, rags, test chemicals, gaskets, paint, etc. Unless demands for any of these items are excessively high, from 0.5% to 1.0% of the plant machinery and equipment cost will be adequate.

Clothing and Laundry. An allowance of one dollar a week for each man engaged in operating labor, including foremen, will be sufficient for average

operations. Special requirements may call for a somewhat higher allowance.

Laboratory. An estimate is made of the amount of analytical work that may be required and the time needed to complete this work. The average salary of an analyst for the time required plus 100% for overhead will be a satisfactory charge in a preliminary estimate.

Royalty. If it is known that the process is wholly or partly covered by existing patents some idea of the probable royalty payments will have to be obtained from the responsible authorities in the company.

Indirect Conversion Expense

Depreciation, M & E. Depreciation rates are fixed by the Bureau of Internal Revenue. For preliminary estimates a rate equal to 10% of the machinery and equipment may be used. This is somewhat higher than the actual average depreciation rates.

Depreciation, Building. Buildings may be depreciated at the rate of approximately 3% of the installed cost.

Taxes. In lieu of actual tax rates, taxes may be taken at 2% of the installed cost of the production machinery, equipment, and buildings for preliminary estimates.

Insurance. An approximate rate for insurance is 0.2% of the total cost of the production machinery, equipment and buildings. Insurance and taxes are sometimes grouped as noncontrollable indirect conversion expenses.

Controllable Indirect Conversion Expense. The controllable indirect conversion expenses include items of factory office and general plant expense such as employee service facilities, watchmen, dispensary, employment office, cafeteria, yard expense and maintenance, purchasing and traffic expenses, etc. To estimate this item accurately specific formulas applicable to the plant location must be obtained from the accounting department.

For preliminary estimates, the controllable indirect conversion expense can be approximated by either (1) 40%-60% of the direct labor, and (2) 15%-30% of the direct conversion expense. Which of the two procedures will be used in a particular case depends upon the judgment of the estimator.

Bulk Conversion Cost. The bulk conversion cost is the sum of the direct and indirect conversion expenses.

Bulk Manufacturing Cost. The net raw material cost plus bulk conversion cost is equal to bulk manufacturing cost. The bulk manufacturing cost is the cost of the end product in storage bins or tanks.

Packaging and Shipping. It is necessary to add the cost of packaging and shipping the product if it is desired to develop an f.o.b. cost.

Containers. Containers mean tank cars, steel drums and cans, bags, boxes, fiber drums and glass vessels. Cost of containers may be secured locally from distributors and establishments selling them.

Tank car rentals are included in this item. In general, tank car rental rates fall within the range $40 to $150, a car a month. A car carrying sulfuric acid would lease for about $56 while one transporting chlorine would have a rental of $150 a month. Special cars such as those made from solid stainless steel with special fittings may rent for as much as $250 a car a month. A new tank car costing $10,000 would rent for $140 to $150 a month for a minimum period of five to seven years.

Packing Labor. Packing labor includes labor involved in transferring the product from bulk storage to shipping containers.

Shipping Supplies and Labor. This item includes the miscellaneous supplies necessary for wrapping, sealing and addressing containers and the labor of loading cars or trucks.

Freight Allowances. Most products are sold f.o.b. the manufacturing plant. If the product is to be delivered to one or more locations for a given price, i.e., freight to be paid by the manufacturer to the point of destination, then the cost of freight must be included with manufacturing and shipping costs.

Freight rates must be obtained for each product since they depend on several factors including distance traveled, nature of the product, hazards in handling, and value of the commodity.

Total Manufacturing Cost, f.o.b. This is the total cost of manufacturing and

delivering the product to the point designated.

Interpretation of Manufacturing Cost Estimate

An analysis of the bulk manufacturing cost estimate (Table 1) will show that only costs entering into actual production are taken into account. Obviously, other important costs such as selling and research expenses, income taxes, and a return on the total capital investment have not yet been considered.

In spite of the fact that the manufacturing cost estimate is an incomplete picture of the project, preliminary economic evaluations have often stopped there. Competing processes have been compared merely on the basis of their manufacturing cost estimates. Experience has shown that this procedure frequently leads to inaccurate conclusions about the economic possibilities of a project that are not discovered until a detailed analysis is made at a later date and after a considerable sum of money has been spent.

In Table 3 two processes for the manufacture of a chemical are compared on the basis of their manufacturing cost and selling price. It will be seen that process A has the advantage of a lower production cost by approximately 2 cents/lb. However, when the capital requirements and other costs are fully evaluated and a selling price is developed, process B has the advantage. The product made according to this process will yield a higher return on a total investment and it can be sold 0.7 cents/lb. less to give the same return as obtained via process A. Process B also has the advantage that it requires only about one-half the total capital investment of process A.

Kinckiner (5) shows the importance of considering the investment in nonmanufacturing and working capital along with the manufacturing capital. He cites three cases in which the return, based on the manufacturing capital only, is 20% in each case. However, when the nonmanufacturing and working capital is added to each, the return on the total investment drops from a uniform 20% to 16, 10, and 7% respectively for the three cases. In the latter two cases the decidedly lower return on the total investment compared with the 20%

return on the manufacturing capital only is due to the greater share of total investment required for nonmanufacturing and working capital.

It is necessary, therefore, to estimate, in the preliminary evaluation at the research stage, all the costs that will be incurred in bringing the product to the ultimate consumer. In the following two sections the procedure for estimating these costs and for developing a selling price will be discussed. The first step is the estimation of the total capital requirements. The capital for M & E and manufacturing buildings already estimated is a part of this total.

Estimation of Capital Requirements

The total capital investment is made up of two kinds of capital: fixed capital and working capital. Fixed capital refers to the capital invested in the permanent manufacturing and nonmanufacturing facilities directly connected with the production of the material. Working capital is the "fluid" capital required for the conduct of the business. Various items that make up the fixed and working capital are shown in Table 4. These items will be discussed in order as follows:

Fixed Capital

Land. Land includes the space occupied by the manufacturing buildings, warehouses, roads, railroad sidings, etc. It means land with improvements. If no value for the land to be occupied is available, $5,000/acre may be used in preliminary estimates.

Buildings. The term "buildings" is ordinarily reserved for the buildings directly connected with the manufacturing operations. However, if new buildings are required for nonmanufacturing or service operations they are included.

Methods for estimating building costs were discussed under the preparation of production cost estimates.

Manufacturing M & E. The manufacturing M & E is the capital devoted exclusively to the manufacturing machinery and equipment. Estimation of this item was summarized in Table 2.

Nonmanufacturing M & E. The nonmanufacturing or service M & E is the capital devoted to the plant functions which serve the direct manufacturing operations. Nonmanufacturing M & E items include installations for genera-

TABLE 3.—TOTAL MANUFACTURING COST VS. SELLING PRICE

	Process A	Process B	
Total manufacturing cost, cents/lb.	18.4	20.3	
Total fixed plus working capital	$5,827,000	2,703,000	
Assumed selling price, cents/lb.	25.7	25.0	25.7
New earnings, dollars	588,000	372,900	351,000
Annual return, after taxes, on total fixed plus working capital	10.0%	10.1%	13.0

tion or distribution of utilities (steam, electricity, water, air, gas); shops, warehouses, and transportation facilities; employee and office facilities; research and control laboratories; and miscellaneous items such as fences, railroads, roads, yard lighting and telephones.

If a plant is to be built in a new location where none of these items of service M & E is available, the nonmanufacturing M & E required must be estimated in the same manner as the items of manufacturing M & E. If the plant is an addition to existing manufacturing operations, only those items of nonmanufacturing M & E not already available must be estimated and recorded as new nonmanufacturing M & E.

General Service Facilities. This category applies to the proportionate share of the existing installations of nonmanufacturing M & E which will be utilized by the new manufacturing plant. Most manufacturing plants are additions to existing manufacturing operations, however, if the manufacturing plant is to be located where no other manufacturing operations are in existence, the general service facilities items will not apply.

Accounting departments develop their own bases for allocating the proportionate share of nonmanufacturing facilities used by the various manufacturing operations. The allocating procedure suggested below is offered so that those who do not have access to specific accounting procedures can estimate these items in an approximate manner. Estimation of nonmanufacturing M & E by allocation is a difficult procedure. The utility of fixed rules is often open to question.

Steam. A basis for assigning the proportionate share of investment in steam generation and distribution facilities used, is to take $5 to $10/lb. of steam used in one hour under average operating conditions.

Power. The share of electric power generation and distribution facilities used may be estimated roughly by taking $100 to $175/kw. hr. used in one hour under average operating conditions. For the distribution system only, $40 to $60/kw. hr. used in an average operating hour may be taken.

Others. The proportionate share of investment in other items of nonmanufacturing M & E, exclusive of steam and power, may be approximated by taking $30 to $50 for each $100 of manufacturing M & E. Although the

majority of cases will be within this range, this rule should be applied cautiously.

Working Capital

Raw Material Inventory. The quantity of a raw material that needs to be held in inventory will vary with each raw material. For a preliminary estimate a one-month raw material inventory may be assumed.

In-process and Finished Goods Inventory. For preliminary estimates in-process goods may be taken at the raw material cost plus one-half the normal conversion cost. Finished goods inventory may be taken at one week's production and priced at the manufacturing cost.

Others. Capital is required to cover credit extended to customers according to the terms of the sale. The terms are usually based on thirty days and capital equivalent to one month's production at cost may be reserved.

Additional cash is required to pay wages and salaries and to purchase raw materials and pay for other operating expenses. Cash equal to the value of one month's production at cost ordinarily will be sufficient for this.

To provide for these two requirements, 15% of the net annual sales may be set aside.

Selling Price and Return on Capital Investment

Two factors required for the estimation of the selling price, i.e., total manufacturing cost and the total investment in capital, have been developed in the preceding sections. Remaining factors to be estimated are:

1. Other expenses not already accounted for, such as cost of research, administration, and selling.

2. Federal and state income taxes.

3. Desired rate of return on total capital investment.

In practice, the selling price is calculated to give a certain rate of return on the total capital investment and takes into account all elements of cost to the manufacturer.

A summary of the procedure for estimating selling price and return on the capital investment is shown in Table 5. A discussion of the items in this table follows:

Assumed Selling Price. A selling price is first assumed and the capital return is then obtained on the basis of income derived from sales at this price by following the steps in Table 5. A trial-and-error procedure may be used to determine a selling price to give a definite return, say 10%, on the total capital investment. If three or more selling prices are assumed and the corresponding return on capital investment calculated in each case, a plot of selling price versus return on investment can be prepared. The selling price at the desired rate of return may then be read from this plot. The estimator may also develop his own formula or procedure for calculating the selling price for a definite return on investment.

Space is allotted in Table 5 for two assumed selling prices to give, respectively, two desired annual rates of return on the capital investment.

Received from Annual Sales, Net. This is the sum realized from the sale of the total annual production of the product at the selling price assumed, less discounts for cash, if any.

Deduct—Annual Production Cost. Total annual manufacturing cost (developed in Table 1), including packaging and shipping costs, and freight allowances, if any, is subtracted from the net annual sales to give the gross profit.

Deduct—Other Expenses, Selling, Administrative. These expenses are those connected with the selling, research, and administrative functions in the company. Selling expense applies to all expenses involved in the sale of the product including advertising. Research expense covers the salaries and overhead for all technical personnel engaged in research and development work. Administrative expenses apply to the salaries and expenses of the officers of the company and to general expenses connected with company administration such as legal fees, auditing fees, etc.

These expenses, lumped together, may be estimated by taking a percentage of the net annual sales. From 5% to 15% of the net annual sales of the product ordinarily will take care of these expenses, the higher figure prevailing where the research budget and selling expenses are relatively high. These expenses subtracted from the gross profit give the net profit. The net profit or operating profit is the income earned before income taxes are deducted.

Deduct—Income and Excess Profit Taxes. The present federal income tax rate is 38% for net corporate incomes more than $50,000 (3). Where state

TABLE 4—ESTIMATED CAPITAL REQUIREMENTS

Product ....X................   Pounds a year 5,000,000

Fixed Capital	
Land	$ 7,000
Buildings	50,000
Manufacturing M & E	392,000
Nonmanufacturing M & E	
General service facilities:	
Steam	32,000
Power	5,600
Others	117,700
Total fixed capital	$604,300
Working Capital	
Raw material inventory	$ 66,000
In process and finished goods inventory	20,100
Others	240,000
Total working capital	$326,100
Total fixed plus working capital	$930,400

TABLE 5—ESTIMATED SELLING PRICE AND RETURN ON CAPITAL

Product ....X................

Annual sales 5,000,000 lb.

	$0.271/lb.	$0.306/lb.
Assumed selling price	$0.271/lb.	$0.306/lb.
Received from annual sales, net	$1,355,000	$1,550,000
Deduct—annual production cost	1,047,000	1,047,000
Gross profit	308,000	483,000
Deduct—other expenses, selling, administrative, etc.	162,500	183,500
Net profit—before taxes	145,500	299,500
Deduct—income and excess profit taxes	55,200	114,000
New earnings	90,300	185,500
Total fixed capital	604,300	604,300
Total fixed plus working capital	892,400	930,400

Annual return on capital:

At $0.271/lb. selling price

	Net profit ($145,500) %	New earnings ($90,300) %
On total fixed capital	24	15
On total fixed plus working capital	16	10

At $0.306/lb. selling price

	Net profit ($299,500) %	New earnings ($185,500) %
On total fixed capital	50	31
On total fixed plus working capital	32	20

TABLE 6—OUTLINE OF REPORT

1. Introduction
2. Summary
3. Conclusions
4. Recommendations
5. Basic Data and Assumptions
6. Description of the Projected Process—Block and Line Flow Sheet
7. Production Cost Estimate and Notes
8. Estimated Capital Requirements
9. Estimated Selling Price and Return on Investment
10. Estimate of the Amount of Additional Work Required and Its Minimum Cost

An outline for a report on evaluations at the research stage is suggested in Table 6. The items are

1. Introduction
2. Summary
3. Conclusions
4. Recommendations

The first four headings are common to most reports and have their usual meaning.

5. *Basic Data and Assumptions.* Laboratory data which have been the basis for the evaluation are summarized together with all assumptions made.

6. *Description of Projected Process —Block and Line Flow Sheet.* The process of manufacture being proposed as the result is described. A block and line flow sheet is included with the description.

7. *Production Cost Estimate and Notes.* Table 1, the detailed manufacturing cost estimate, is included with notes on the derivation of some of the items.

8. *Estimated Capital Requirements.* Table 4, summarizing the capital investment needed, is shown. Sometimes it is necessary to distinguish between the new capital required for the purchase of manufacturing and nonmanufacturing machinery and equipment, and capital already existing in items of equipment that are definitely known to be available for manufacture of the new product.

9. *Estimated Selling Price and Return on Investment.* Table 5 showing two selling prices to give, respectively, two rates of return on the capital investment is incorporated in the report.

10. *Estimate of Amount of Additional Work Required and Its Minimum Cost.* An estimate is made of the amount of additional laboratory, bench scale, and pilot plant work needed to supply the missing information. This is done ordinarily in connection with the engineering evaluation of the project.

income taxes apply, they are approximately 5% to 10% of the net corporate income.

New Earnings. New earnings are obtained by subtracting the income taxes from the net profit. They represent the sum of money that has been earned by the over-all investment of capital and labor to make the product available to consumers.

Total Fixed Plus Working Capital. From Table 4.

Annual Return on Capital. The annual return on capital is usually expressed on a percentage basis. Net profit and new earnings are shown as percentages of the fixed capital and of the total fixed plus working capital required.

Provision has been made in Table 5 for calculation of the annual return based on two selling prices. This has been done to show how much the selling price must be increased to give a certain increase in the annual capital return. In the example in Table 5, selling prices were estimated to give returns of 10% and 20% on the total capital investment, based on new earnings.

The actual return that will be required by management for a given manufacturing operation depends upon a number of factors and cannot be predicted at the research stage. Therefore, a selling price may be calculated to give an average rate of return in one case and a considerably higher rate in the second case.

Report on Engineering and Economic Evaluations

A report should be written, presenting pertinent facts and conclusions reached, and distributed to all those who are likely to have any connection with the project.

Since the evaluation based on several assumptions may reflect the personal judgment and bias of the chemical engineer preparing it, all assumptions and supporting experimental data should be clearly stated.

From the extent of these requirements, a rough estimate of the minimum cost to obtain the data can be prepared.

Short Cuts in Estimating

The chemical engineer is frequently asked to give a horseback estimate of production cost and probable selling price of a chemical. The quick estimate is required when it is necessary to make a rough evaluation of a suggestion before any further consideration is given, when time is too short, or when there are insufficient data for the longer procedure. If possible, however, the detailed evaluation should be made because information in addition to the manufacturing cost and approximate selling price may be derived from a more thorough study.

In general, a chemical engineer making preliminary estimates will develop his own system of short cuts. He may, for example, correlate conversion costs, over a long period of time, with the kind and number of unit operations and processes involved. His familiarity with certain operations may permit him to make accurate predictions on the cost of similar operations.

There appears to be no substitute for experience in acquiring facility in the use of short cuts. The following procedure is suggested for those who are relatively inexperienced yet wish to prepare rough preliminary manufacturing cost estimates.

1. From the material balance, determine raw material requirements and annual raw material cost.

2. Sketch a flow sheet of the process, obtain the rate at which the product is to be produced, and determine roughly size and cost of important items of equipment. Multiply total purchase price of equipment by a factor (7), (8), to obtain total capital required for the manufacturing machinery and equipment installed. Use of this factor to obtain the capital needed for manufacturing operations was discussed in the derivation of the items in Table 2. Make a rough estimate of cost of process buildings.

3. Estimate utility requirements and total annual utilities bill.

4. Estimate approximate number of man-hours of labor and annual labor bill.

5. Estimate annual cost of direct supervision.

6. Make a rough estimate of annual laboratory expenses.

7. Multiply annual labor plus supervision cost by a factor of 1.79 to obtain the approximate annual cost of labor plus the items in production cost estimated from labor bill.

8. Multiply capital required for machinery and equipment (Item 2, above) by a factor of 0.202 to obtain annual cost of fixed charges plus repairs.

9. Multiply building cost by 0.052 to obtain annual fixed charges connected with use of buildings.

10. Total annual manufacturing cost is then obtained by adding:

   a. Annual bill for raw materials
   b. Annual utilities bill
   c. Annual laboratory expense (Item 6)
   d. Item 7
   e. Item 8
   f. Item 9

When the procedure outlined in Item (10), above, is applied to the example in Table 1, using the capital for machinery, equipment, and buildings, the annual raw materials cost, etc., the following total annual manufacturing cost is obtained:

a.	$792,460
b.	56,590
c.	3,000
d.	72,100
e.	79,200
f.	2,600
	$1,005,950

$$\frac{\$1,006,000}{5,000,000} = \$0.201/lb.$$

This cost compares with annual bulk manufacturing cost of $1,013,900, equivalent to $0.203/lb., obtained by the detailed procedure in Table 1.

The shorter procedure outlined will be accurate enough for most preliminary estimates. The accuracy may be increased in some cases by the adjustment of the factors in Items (7), (8), and (9), when it is known that special conditions apply which may change the factors appreciably.

A useful rule for estimating the cost of an item of equipment of a given size from the known cost of the same item in a different size is the Six-Tenths Factor discussed by Williams (15). Cost of buildings and complete manufacturing plants may be projected also to other sizes if the cost of these in one size is known.

The selling price is difficult to estimate by any short-cut rule. Relative amounts of nonmanufacturing M & E and working capital compared with the manufacturing M & E, and other factors as well, must be known. These may be estimated by the somewhat abbreviated methods already described in con-

nection with the development of information in Tables 4 and 5.

Conclusions

With the increasing cost of research and process development work leading to the commercial production of chemicals it has become imperative that the engineering problems to be encountered and the economic possibilities of a research project be understood early. Management must know, as soon as possible: (a) how much it may cost to manufacture the product at the desired annual production rate, (b) how much new capital may be required, (c) what net annual return might be realized from the capital invested at the selling price visualized, and (d) if the foregoing analysis shows that the project is an attractive one, the probable amount of process development work involved and the possible cost of this work.

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OXYGEN-ENRICHED AIR

THERMODYNAMIC ANALYSES OF PROCESSES FOR ITS MANUFACTURE

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Processes for the manufacture of enriched air of 30, 40, and 50% O_2 content by rectification of atmospheric air are considered. Minimum energy requirements for reversible separation are shown to be 0.174, 0.320, and 0.447 hp. hr./lb. mole of oxygen contained in the enriched air product for the three cases above. These are approximately half the energy requirements for the dilution of reversibly produced pure oxygen to the above compositions.

When practical processes are considered, the vital importance of the distilling column is pointed out, and the Linde double column is studied with particular regard to the maximum amount of extra atmospheric pressure air which may be provided to utilize the extra reflux available in the upper part of this column. Enthalpy-concentration tables at 30, 40, 50, 60, and 70 lb./sq.in.abs. are developed for this purpose.

Three different processes, all with external purification, are analyzed: the cascade without extra air, a low pressure expander process without extra air, and the same with the maximum possible extra air. The last is, of course, the best and leads to the following energy requirements for practical conditions:

30%	0.06 hp. hr./lb. O_2	(1.92/mole)
40%	0.071 hp. hr./lb. O_2	(2.28/mole)
50%	0.083 hp. hr./lb. O_2	(2.66/mole)

These correspond to low efficiencies (based on reversible separation), and the efficiencies fall as the oxygen contents fall.

MUCH has been written in recent years about tonnage oxygen and its great potentialities for many new developments. Most of the emphasis has been put on a 95% or greater oxygen content, but many attractive applications do not require such purity. For example, in the steel industry, air enriched to only 26% O_2 appears to have much promise, and in sulfuric acid manufacture and nonferrous metallurgy 30 to 60% O_2 would be of value. These and other potential uses of oxygen are thoroughly considered by Downs and Rushton (7).

Oxygen-enriched air may be made in a variety of ways, but the entire attention of this work will be directed to only those processes which depend on low temperature rectification of air. Even with this limitation two alternatives exist: one to make high purity oxygen and dilute it with air to the desired composition and the other to rectify in the column only to the desired oxygen con-

tent. The first of these needs no further examination here, since energy requirements may be estimated from the results of others already available. The second will be the principal subject of this study. It is immediately obvious that the dilution method will involve a considerably smaller amount of material to be handled at the low temperature, and this is an advantage for this method. On the other hand the energy requirement, based on purely reversible separation, is appreciably lower for the direct process. If the comparative advantage of the direct process in energy requirement could be shown to be capable of practical attainment, this advantage is probably great enough to offset the disadvantage of handling more material in the low temperature part of the system. It may be noted that some writers (1, 9, 13) state that the dilution method is better, but these statements are so incomplete that it is by no means certain

that all factors have been considered. Details of the distillation column which must be operated at nearly the minimum reflux for low energy requirements are particularly lacking.

Oxygen-enriched air is not a definite term because of the variation in possible oxygen contents, and in order to facilitate comparisons attention will be limited to 30, 40, and 50% O_2. Methods to be described are, of course, applicable to any desired composition.

In a recent paper Bliss and Dodge (2) reported results of extensive thermodynamic analyses of processes for the manufacture of essentially pure oxygen, and the same viewpoint will be maintained in this paper. These methods lead to much information about operating pressures, energy requirements, and necessary distillation facilities, but they do not lead to specifications of equipment sizes.

It is therefore the purpose of this paper to present methods for and results of thermodynamic analyses of processes for the manufacture of 30, 40, and 50% O_2 by rectification of air and to compare their energy requirements with those to be expected with the dilution of high purity oxygen to these compositions.

Conditions, Assumptions, and Properties

Many conditions and assumptions used by Bliss and Dodge (2) will be used here also. Theirs and others are summarized thus.

Pure air containing 21% O_2 and 79% N_2, free of argon, carbon dioxide, water and hydrocarbons will be assumed as the raw material, i.e., purification is external. The neglect of argon is certainly minor except in the calculation of the number of plates in an actual distillation column, a problem not here considered.

Compressors will be assumed to be 75% efficient based on isentropic compression of an ideal gas, and expanders

will be assumed to be 80% efficient based on isentropic expansion of the real gas being handled.

Pressure drops due to friction will be neglected except in certain low pressure cases when a drop of 5 lb./sq.in. is assumed.

Initial temperature is 80° F. in all cases.

Initial pressure is a variable of considerable importance which will be discussed in more detail later. There are certain minimum values which are treated in the distillation section to follow.

Minimum terminal temperature approaches will be 3° C. In certain cases, particularly the exchange of latent with sensible heat, it is often impossible to fix the terminal temperature approaches since the minimum approach will be within the exchanger and in these cases the figures here are relaxed. Only a finite temperature difference of the right direction to assure the proper direction of heat flow is required in these cases.

Temperature of expander intake is a variable for the processes here treated and the fraction of gas expanded is also a variable, nearly always dependent.

Purity of the product nitrogen is 100% in all the cases here considered.

Heat leak has been assumed to be 28 B.t.u./mole of air treated.

Thermodynamic properties were read from the charts and tables of Williams (14) and Lobo (10) for air and Millar and Sullivan (11) and Lobo (10) for oxygen and nitrogen. The vapor-liquid equilibirum data of Dodge and Dunbar (6), and the enthalpy concentration information of Ruhemann (12) and Lobo were used. New enthalpy concentration diagrams, discussed below, were constructed for this work.

Classification of Processes

In the work of Bliss and Dodge (2) processes were classified according to the manner of accomplishing refrigeration as follows:

1. Joule-Thomson effect only
2. Joule-Thomson effect plus auxiliary liquid-vapor refrigeration at moderately high temperature levels relative to that of liquid air
3. Joule-Thomson effect plus approximately reversible expansion of air or products in an expander
4. Refrigeration essentially due only to approximately reversible expansions of auxiliary fluid or fluids such as helium through expanders
5. "Cascade" processes
6. Refrigeration essentially due only to approximately reversible expansion of air or products in an expander.

TABLE 1—MINIMUM WORK REQUIREMENTS FOR REVERSIBLE SEPARATION OF AIR INTO GASEOUS NITROGEN AND OXYGEN-ENRICHED AIR

Ambient Temperatures, 300° K.

x_2 Mole Fraction of Oxygen in Enriched Air	x_3 Mole Fraction of Oxygen in Waste	W, hp.hr./lb. Mole of Pure Oxygen in Enriched Air
0.3	0	0.174
0.4	0	0.320
0.5	0	0.447
0.7	0	0.663
1.0	0	1.035
0.3	0.01	0.152
0.4	0.01	0.293
0.5	0.01	0.408
0.7	0.01	0.615

In the case of enriched air, however, and where low energy requirement is of primary interest, processes 1 and 4 must be excluded because they were so lacking in promise for pure oxygen. Process 2 is not possible here because the pressures are so low that the heat capacities of incoming air and returning oxygen and nitrogen are very nearly the same. Process 3 is practically indistinguishable from process 6 at these low pressures. Only numbers 5 and 6 are left, and all attention in this work will be directed to them.

Bliss and Dodge (2) used the character I to denote liquid oxygen processes and II to designate gaseous ones. Here III will be used for enriched air with subsequent letters to denote a special subcase of process 5 or 6.

Theoretically Minimum Energy Requirements

As Bliss and Dodge (2) have shown the minimum energy requirement for reversible separation may be computed by the equation

$$W = \Delta H - T_o \Delta S = -2.3 R T_o$$
$$\{M_2[x_2 \log x_2 + (1 - x_2)$$
$$\log (1 - x_2)] + M_3[x_3 \log x_3$$
$$+ (1 - x_3) \log (1 - x_3)]$$
$$- M_1[0.21 \log 0.21 + 0.79 \log 0.79]\}$$
$$(1)$$

in which M_1, M_2, and M_3 are the moles of air, enriched-air product, and nitrogen product and x_2 and x_3 are the oxygen contents of the enriched air and nitrogen products. It was there shown that if x_2 or x_3 become 0 or 1 in the above equation the term containing this value vanishes. Figures for the three cases to be considered and for a few others are shown in Table 1.

These figures will be used as bases for efficiency figures to be computed later. Simple material balances will show that the first three figures in Table 1 would become: 0.393, 0.621, and 0.758 hp. hr./ lb. mole for the dilution process, that is for the dilution of reversibly produced pure oxygen. These figures are clearly

in the neighborhood of twice those for the direct process, so the theoretical advantage of the direct process from an energy requirement standpoint is apparent. Whether this theoretical advantage can be practically attained remains to be seen.

Enthalpy-Concentration Diagrams

Since enriched-air processes may be operated at pressures below those for manufacture of pure oxygen it is necessary to have enthalpy-concentration diagram between 1 and 5 atm. at which pressures such information was already available. Accordingly, several new diagrams have been prepared for 30, 40, 50, 60, and 70 lb./sq.in. abs. It would be desirable to have such diagrams for even closer pressure intervals, but these were sufficient for the cases considered. The method of construction for any one pressure is delineated below.

Values of the enthalpies of saturated vapor and liquid oxygen and nitrogen were interpolated from the data of Millar and Sullivan and plotted as terminal points on the desired H-x diagram.

Values of the enthalpies of superheated nitrogen and oxygen up to 150° K. were determined from the same source and plotted. Straight lines (assuming no heat of mixing in the superheated region) were drawn between the pure oxygen and nitrogen points at the same temperature. It was observed that the slopes of these isothermal lines were dependent on temperature and in fact that the change in slope per degree Kelvin was approximately constant. This observation made it possible to construct these straight lines for temperatures at which nitrogen was still a superheated vapor but at which oxygen could not exist as a vapor, that is the lines were determined by the nitrogen terminus and the slope rather than the nitrogen and oxygen termini since the latter cannot be determined in this range. These lines actually represent physical reality only in that range of x between pure nitrogen and the saturated vapor line. The pressure of the diagram and the temperature of the line of course define the dew-point composition which is given in the Kellogg Charts (10). This composition is then a point on the saturated vapor line. A series of such points was thus determined and they were then connected to establish the saturated vapor line.

The saturated liquid line was located by subtracting latent heat of vaporization at constant composition from enthalpy on the saturated vapor line. Latent heats were determined as: Latent heats of air, oxygen and nitrogen for pressures within the range of 14.7 to 75

lb./sq.in.abs., were determined by interpolation of results of Williams (14) and Millar and Sullivan (11). These were then plotted vs. pressure for the three compositions and a straight line drawn between 14.7 and 75 lb./sq.in.abs. Slopes of these straight lines were plotted vs. composition and the deviations of the interpolated latent heats from these straight lines at the desired pressure were also plotted vs. composition for the three cases: air, oxygen and nitrogen. Also a plot of Dana's (4) data on the heat of vaporization at atmospheric pressure vs. composition was made. Using the plot of Dana's data it was then possible to determine the latent heat of any desired mixture at 1 atm. The effect of pressure on this latent heat was then determined from the two plots described above, one which might be called a linear effect and the other deviations from the linear effect. Latent heats determined in this way were then subtracted from the saturated vapor line to give the saturated liquid line.

Tie-lines to express vapor-liquid equilibrium were determined from the equation

$$\frac{y}{1-y} = \frac{ax}{1-x} \qquad (2)$$

the values of a being interpolated from the data of Dodge (5) to yield

$a = 0.296$ at 30 lb./sq.in.abs.

$= 0.320$ at 40 lb./sq.in.abs.

$= 0.340$ at 50 lb./sq.in.abs.

$= 0.358$ at 60 lb./sq.in.abs.

$= 0.374$ at 70 lb./sq.in.abs.

a is defined by Equation (2) but it should be noted that these are mole fractions of oxygen since a has values less than 1.0.

Rather than present this information in figures these results are presented in Tables 2 and 3 so that they may be plotted to the scale desired.

It should be noted that these tables are presented in terms of mole fractions of nitrogen which is the more volatile component.

Rectification Considerations

Since our point of emphasis is low energy requirement only the best types of columns will be considered. The simple column, since it always produces a low yield of oxygen, is thus excluded. Another type of column, one properly referred to as an enriching column and which operates under full process pressure, has been suggested. Feed air enters below the plates and the enriched-air product as a liquid is withdrawn, throttled, and evaporated in condenser coils to provide refrigeration for reflux at the top of the column. This device

is capable of producing enriched air of 30 and 40% O_2 content but cannot yield 50% because of equilibrium considerations at the bottom of the column. It is discarded from further consideration here primarily because it has no capacity for handling extra low pressure air (since it operates under full process pressure), and this extra air is all important in reducing energy requirement.

Actually the whole attention in this section is devoted to the double column, although it should be noted that the compound column with reflux from an auxiliary nitrogen circuit is probably of comparable performance. It is well known that the 1-atm. section of the double column has considerable excess reflux, and since interest here is primarily in low energy requirement it will be found desirable to utilize this excess reflux. This can be done most readily by feeding to the double column some extra air (the Lachmann principle), from an auxiliary circuit which extra air costs little in terms of energy requirement. It will always be assumed that this extra air is saturated vapor at 1 atm.

It is the primary aim of this section to develop the methods by which the maximum permissible amount of extra air may be computed and the results of the calculations thereon.

In this paper discussion is limited to the three cases of 30, 40, and 50% O_2, although it will be helpful in this section to think of them as 70, 60, and 50% N_2.

Minimum Pressures. A 3° K. temperature difference between the condensing nitrogen and boiling enriched air will be assumed. It is assumed and later confirmed that it is possible to rectify to pure nitrogen in the lower (higher pressure) column. The minimum pressures thus determined are
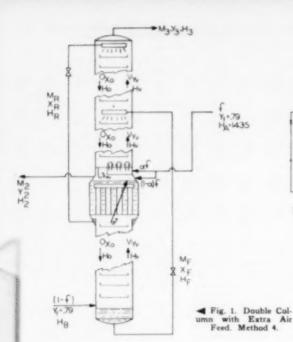
30% O_2........36.5 lb./sq.in.abs.

40% O_2........41.5 lb./sq.in.abs.

50% O_2........46.5 lb./sq.in.abs.

To simplify certain calculations in the study of permissible extra air, pressures of 40 (for 30% O_2) and 50 lb./sq.in. (for 40 and 50% O_2) will be used for the lower column, since it is necessary to examine the lower column on an H-x diagram and it is impractical to construct them for more than 10 lb./sq.in. intervals.
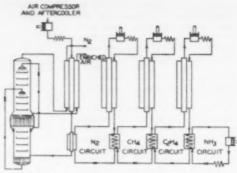
Extra Air Feed Location. In the case of pure oxygen the proper feed location for the extra air is easy to establish, since it is simply that plate over which the vapor composition is 79% N_2. In the case of enriched air over the range here considered (70, 60, and 50% N_2), there

(C.h.u./lb.mole)

Mole Fraction N₂	0	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9	1.0
					$P = 30$ lb./sq.in.abs.						
H_v	1805	1770	1729	1688	1645	1605	1561	1515	1470	1425	1377
ΔH	1555	1549	1535	1513	1487	1459	1425	1391	1355	1315	1276
H_L	253	221	194	175	158	146	136	124	115	110	101
					$P = 40$ lb./sq.in.abs.						
H_v	1819	1792	1743	1702	1660	1616	1575	1529	1487	1439	1391
ΔH	1522	1515	1502	1480	1455	1427	1396	1362	1326	1296	1246
H_L	297	267	241	222	205	191	179	167	161	153	145
					$P = 50$ lb./sq.in.abs.						
H_v	1828	1792	1751	1711	1671	1630	1584	1540	1495	1449	1400
ΔH	1493	1486	1475	1453	1428	1400	1368	1336	1299	1258	1218
H_L	335	306	276	258	243	230	216	204	196	191	182
					$P = 60$ lb./sq.in.abs.						
H_v	1834	1797	1756	1718	1676	1635	1591	1547	1503	1455	1405
ΔH	1466	1460	1448	1427	1402	1376	1343	1310	1274	1234	1193
H_L	368	337	310	291	274	259	248	237	229	221	212
					$P = 70$ lb./sq.in.abs.						
H_v	1837	1803	1764	1724	1682	1639	1594	1551	1508	1461	1409
ΔH	1442	1435	1426	1406	1381	1354	1323	1280	1253	1211	1170
H_L	395	368	338	318	301	285	271	262	255	250	239

TABLE 3.—VAPOR-LIQUID EQUILIBRIUM

Oxygen-Nitrogen

Interpolated from Dodge (5)

Mole Fraction N₂ in Liquid	Mole Fraction N₂ in Vapor				
	$P = 30$ lb./sq.in.abs. $a = 0.296$	$P = 40$ lb./sq.in.abs. $a = 0.320$	$P = 50$ lb./sq.in.abs. $a = 0.340$	$P = 60$ lb./sq.in.abs. $a = 0.356$	$P = 70$ lb./sq.in.abs. $a = 0.374$
0.1	0.273	0.257	0.246	0.237	0.229
0.2	0.458	0.439	0.424	0.411	0.401
0.3	0.592	0.572	0.558	0.545	0.534
0.4	0.693	0.675	0.662	0.651	0.641
0.5	0.772	0.757	0.746	0.737	0.728
0.6	0.845	0.824	0.815	0.807	0.800
0.7	0.887	0.879	0.873	0.867	0.862
0.8	0.931	0.926	0.922	0.918	0.914
0.9	0.968	0.966	0.964	0.962	0.960

$$M_3, y_3, H_3$$

$$f$$
$$y_1 = .79$$
$$H_A = 1435$$

$$M_R$$
$$X_R$$
$$H_R$$

$$M_2$$
$$y_2$$
$$H_2$$

$$\alpha f$$

$$(1-\alpha)f$$

$$M_F$$
$$X_F$$
$$H_F$$

$$(1-f)$$
$$y_1 = .79$$
$$H_B$$

▲ Fig. 2.
Process 5-III.

is no exactly appropriate place for the extra air feed. Several places may be considered as follows:

1. The extra air may be mixed with the oxygen-rich product of the lower column and become the feed to upper column.

2. It may be fed to the still pot of upper column.

3. A part of it may be fed to upper column still pot and a part of it to the vapor leading to column proper, the oxygen-rich product from lower column being also fed into the pot.

4. Part of it may be fed to upper column still pot and a part of it to the vapor leading to the column proper, the oxygen-rich product from the lower column being sent to a plate in upper column.

Method 4 is illustrated in Figure 1 and seems to be the best in all cases, the best being defined in terms of permitting the greatest quantity of extra air. It is interesting to note that method 1 is practically as good as 4 and superior to 2 and 3 for 50% O₂, and that method 3 is nearly as good as 4 for 30% O₂.

Method and Calculations for 70% Nitrogen Case. Operating pressure in lower column is 40 lb./sq.in.abs. The

first law requires that the column as a whole (Fig. 1) satisfy these equations

$$1 = M_2 + M_3$$

$$0.79 = y_2 M_2 + y_3 M_3 \qquad (4)$$

$$(1-f)H_B + fH_A = M_2 H_2 + M_3 H_3 \qquad (5)$$

neglecting heat leak into the column.
For this particular case $y_2 = 0.7$, $y_3 = 1.0$, $H_A = 1435$, $H_2 = 1475$, $H_3 = 1335$ whence

$$M_2 = 0.7$$

$$M_3 = 0.3$$

$$H_B = \frac{1435 - 1435f}{1-f} = 1435$$

Turning attention to the lower column the first law requires that

$$(1-f) = M_F + M_B \qquad (6)$$

$$(1-f)0.79 = M_F x_F + M_B x_B \qquad (7)$$

$$(1-f)H_B = M_F H_F + M_B H_B + Q_{CB} \qquad (8)$$

In order to solve these equations it is necessary to take trial values of x_F and x_B. The value of x_F may be anywhere between 0.79 and 0.57 (limited by equilibrium with the feed to the lower column). Experience has shown the lowest permissible value to be best. Assuming $x_F = 0.57$ and $x_B = 1.0$ (later con-

firmed). $H_F = 180$ and $H_B = 148$. Solving these equations we have

$$M_F = 0.488(1 - f)$$

$$M_B = 0.512(1 - f)$$

$$Q_{CB} = 1271(1 - f)$$

The Ponchon-Savarit operating lines for the lower column may be shown to be defined by

$$\frac{x_B - x_o}{Q'' - H_o} = \frac{y_v - x_o}{H_v - H_o} \qquad (9)$$

in which

$$Q'' = H_B + \frac{Q_{CB}}{M_B} \qquad (10)$$

Therefore

$$Q'' = 148 + \frac{1271(1-f)}{0.512(1-f)} = 2628$$

It is notable that this is independent of f, a result found to be true in all cases studied. The tie-line through the lower column feed extended to $x_B = 1.0$ cuts the latter at a point well below 2628. The lower column is therefore operable with ease regardless of the value of f and capable of yielding $x_B = 1.0$ as assumed.

Turning attention to the upper part of the upper column, the Ponchon-Savarit operating lines are

$$\frac{y_3 - y_v}{Q_T - H_v} = \frac{x_o - y_v}{H_o - H_v} \qquad (11)$$

in which

$$Q_T = \frac{M_3 H_3 - M_B H_B}{M_3 - M_B} \qquad (12)$$

Equation (12) may be used to determine the value of f directly by extending the tie-line through the upper feed, $x_F = 0.57$, $H_F = 180$ (on a 1-atm. H-x diagram) to the intersection with $x_B = 1.0$ corresponding to the use of an infinite number of plates above the feed. This intersection, the limiting value of Q_T, is 2200.

Therefore

$$2200 = \frac{3 \times 1335 - .512(1-f)148}{.3 - .512(1-f)}$$

by which

$$(1 - f) = 0.247$$

and, rounding off,

$$f = 0.75$$

This fraction of extra air is operable in the top part of the column, but it remains to be shown that it is in the lower part.

Balance equations for the whole upper column excluding the pot are

$$M_F + M_R + af = M_3 + O_C \tag{13}$$

$$M_F x_F + M_R x_R + af x_1 = M_3 y_2 + O_C x_C \tag{14}$$

$$M_F H_F + M_R H_R + af H_4 = M_3 H_3 + O_C H_C \tag{15}$$

It will be noted that it has been assumed that no vapor to the upper column comes from the still pot. Since this vapor is the leanest in nitrogen of all vapors in the region this is proper. The unknowns in the above equations are O_C, x_C, and a, since H_C is defined by x_C. Solution of the above equations leads to these results

a (fraction of extra air fed to upper
　　column as distinct from pot)
　　　　　= 0.345
　　　　$O_C = 0.209$
　　　　$x_C = 0.49$
　　　　$H_C = 60$

It is now necessary to establish that this will work, i.e., that $x_C = 0.49$ is a permissible value. It seems clear that x_C cannot be less than that value in equilibrium with $y_1 = 0.79$ for the extra air, which value is 0.48. Thus operability is assured. As a double check the Ponchon-Savarit operating lines for the section below the feed have been derived, and it has been shown that the operating point thus defined is such that the lower part of the upper column is operable to yield $x_C = 0.49$.

It has also been established that other values of x_F lead to lower values of f. Therefore we may conclude that this column is operable to yield 70% N_2 and 100% N_2 products when 75% of the air fed to the whole assembly is actually fed as 1 atm. saturated vapor to the upper column. A portion, 34.5%, of this 75% should be fed to the upper column proper and the rest to the still pot.

Similar calculations, for 60 and 50% N_2 lead to results which are shown in Table 4.

These values will be used in a later section in devising the best processes, and it will be assumed that the permissible extra air is unchanged by operation of the process at the minimum pressure instead of the slightly higher values above.

Process Analyses

Methods for analysis of the various processes are the same as described by Bliss and Dodge (2) for pure oxygen, and they will not be described in detail here. One merely must apply the first law to the process as a whole and to each unit thereof and solve for the various unknowns. Heat exchangers in particular are then examined in the light of the second law to be sure that no virtual violations occur.

Only processes 5 and 6 have been studied, and the results are shown in Table 5. The process description and the results are discussed below. The effects of such variables as heat leak, return gas temperature, etc., have not been examined here, since they should have about the same effects as with pure oxygen.

Process 5-III. This process, using four refrigerants, is illustrated in Figure 2. A return gas temperature of 297° K., a heat leak of 28 B.t.u./lb. mole of air, and the use of a double column were assumed. In addition it was assumed that all the heat leak was concentrated in the heat exchanger where the refrigeration with nitrogen was performed, and it was further assumed that the pressure of operation was the minimum for each case as established in a previous section. Extra air was not provided in these cases.

It is clear, from Table 5, that the cascade has large energy requirements compared with the reversible process. It is also worthy of note that the computed energy requirement rises with oxygen content of the product much less than the reversible energy requirement, i.e., the efficiency falls as oxygen content is decreased. Thus the efficiency is 10.5% for 50% O_2, 8.2% for 40%, and 4.99% for 30% O_2. This is probably in large part due to the greater excess reflux in the column in the low purity cases, i.e., greater irreversibility.

Process 6-III. This is illustrated in Figure 3. Again the return gas temperature was 297° K., the heat leak 28 B.t.u./lb. mole air, and the double column used. In this case, the expander
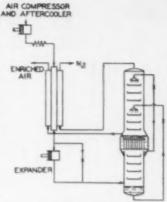
Fig. 3. Process 6-III.

exhaust was set at the minimum pressure of operation of the column, which, of course, requires that the initial pressure be somewhat higher.

It is clear from these results that this process is comparable with the cascade, and in fact the differences in energy requirement between the two are probably insignificant. The same small increase of energy requirement with product oxygen content (i.e., decreased efficiency with decreased oxygen content) is observed and presumably for the same reasons.

Process 6-IIIA. This process (Fig. 4) differs from 6-III only in that excess air at 1 atm. has been provided in an amount equal to the theoretical maximum as previously determined. Thus 75% of the total air was at 1 atm. for 30% O_2, 59 for 40%, and 49 for 50%. It should be noted that 5 lb./sq.in. pressure drop has been allowed for the low pressure air heat exchanger. This has not been done for other exchangers because of the appreciably higher pressure, but it seemed to be rather important to allow this much here, since the pressure is so low. It has been assumed that the permissible extra air figures cited above are not changed by the slight change in pressure from that of the H-x diagram to the minimum for operation.

Results show considerable improvement over 6-III for all cases. Efficiency of the processes is again lower for the lower oxygen purity cases. This is not explainable by excess reflux, because it

TABLE 4.—RESULTS OF CALCULATIONS ON PERMISSIBLE EXTRA AIR

Oxygen Content Enriched Air %	Pressure of Lower Column lb./ sq.in.abs.	x_C Mole Fraction Nitrogen	f (Fraction of Total Air Which May Be Fed At 1 Atm.)	a (Fraction of Extra Air Fed to Column Proper)
30	40	0.57	0.75	0.345
40	50	0.58	0.59	0.68
50	50	0.58	0.49	0.99+

Double Column, Entrance Air at 300° K. Return

Gas Temperature 293° K., Heat Leak 28 B.t.u./Mole Air

Waste Pure Nitrogen, Condenser-boiler ΔT 3° K. in All Cases

Process	30% Oxygen		40% Oxygen		50% Oxygen	
	5-III	6-III-A*	5-III	6-III-A*	5-III	6-III-A*
Entrance air pressure, atm	2.48	3.2	2.82	3.61	3.16	4.01
Nitrogen rate, lb. mole air	1.1		1.06		1.02	
Nitrogen intake and discharge pressures, atm	2.6,14.5		3.2,19.5		3.9,18.5	
Methane intake, lb. mole air	0.51		0.49		0.48	
Methane intake and discharge pressures, atm	1.24,8		1.24,8		1.24,8	
Ethylene rate, lb. mole air	0.45		0.44		0.42	
Ethylene intake and discharge pressures, atm	1.19.5		1.19.5		1.19.5	
Ammonia rate, lb. mole air	0.11		0.11		0.10	
Ammonia intake and discharge pressure, atm	1.11.5		1.11.5		1.11.5	
Expander inlet temperature, °K	94	93(26%),74(26%)	93.6	94.4(1%),86(59%)	96.8	96(51%),86(49%)
Expander inlet and outlet pressures, atm	3.37,48	3.32,2.48(26%),1.29.1	3.61,62	3.32,62.4(1%),1.29.1(59%)	4.0,53.36	3.74,3.16(51%),1.29.1(49%)
Portion of air expanded, %	90	100	90	100	90	100
Energy requirement, hp. hr./lb. O2	0.109	0.060	0.125	.071	0.133	0.083
Efficiency, per cent†	4.99	9.65	6.72	7.75	7.4	20.5

* In these cases only extra air was provided in the theoretically maximum amount. Pressure drop of 5 lb./sq.in. allowed for heat exchanger in low pressure stream. Heat leak concentrated in exchangers and apportioned between the two streams approximately as their amounts.

† Based on reversible separation.

‡ This figure is approximate, since it is difficult to read thermodynamic diagrams with accuracy in this region.

all these cases the reflux ratio is at the minimum in one part of the column. The basic reason for the poorer performance of the lower oxygen cases probably lies in the condenser-boiler of the double column. The necessary pressure (defined by the temperature difference of 3° K. across the condenser boiler) for 30% O_2 is lower than that for 40 and 50%, but not enough lower to reduce to any great degree the work of compression. Furthermore, the lesser the oxygen content the more nitrogen must be boiled in the condenser-boiler, and heat transfer here is irreversibly accomplished.

Conclusions

Since Bliss and Dodge (2) have shown that pure oxygen could be obtained with the best process for about 0.14 hp. hr./lb. simple material balances indicate that 30, 40, and 50% O_2 could be made by dilution of high purity oxygen with air for

30%	0.053 hp.hr./lb. oxygen
40%	0.084 hp.hr./lb. oxygen
50%	0.103 hp.hr./lb. oxygen

The direct process is clearly a little poorer for 30%, but it is somewhat better for 40 and 50% O_2. It is so little better, however, that it is doubtful that it would be used in view of the greater amount of purification required. Hochgesand (8) gives a figure for 40% O_2 almost the same as the above, but the basis of his result is not clear. If one chooses to compare the direct process with dilution on the assumption that pure oxygen requires 0.2 hp. hr./lb., the above figures become

30%	0.076 hp.hr./lb. oxygen
40%	0.12 hp.hr./lb. oxygen
50%	0.147 hp.hr./lb. oxygen

With such a comparison the direct process would appear to have some possible attraction for 40 and 50% cases if the scale of operations is big enough to warrant the added capital cost in order to reduce the power cost. It may be noted that for 50% O_2, only about 36% more material must be handled in the low temperature part of the system of the direct process than in the dilution process.

Returning to the first set of figures which seems to be more fair since best enriched air and best pure-oxygen processes are compared, it is clear that the full theoretical advantages of the direct process (i.e., the much lower reversible work of separation) cannot be realized in practical processes. There are probably two major reasons for this. The first of these is the column. The necessary minimum pressure for operation of the rectifying column is 30 to 50 lb./sq.in.abs. for enriched air and only about 60 lb./sq.in. for pure oxygen. The second is that there is little increase in

work of compression if the discharge pressure is 60 lb./sq.in.abs. instead of 30 to 50. Thus, pure oxygen can be produced for only little more work than is required to produce enriched air, and with that pure oxygen considerably more enriched air can be made by simple dilution and for no additional energy.

Therefore, for enriched air of low oxygen content, e.g., near 30% O_2, it seems clear that dilution of high purity oxygen is definitely preferable. For higher oxygen contents the two alternatives become more competitive, and a more thorough study in the light of local particular requirements would have to be made to reach a decision between the two. In such cases the figures on energy requirements of this paper should be of value.
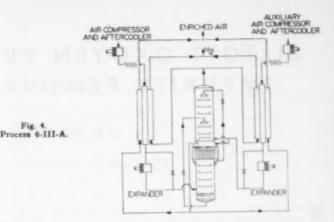
The conclusion that enriched air of low oxygen content should not be produced directly seems to have been independently arrived at by the Linde Co. of Germany, since of seventy-three Linde-Fränkl units listed by Clark (3) only one was built to produce 45% O_2. The rest were to produce 95% O_2 or purer. Bishop (1), Karwat (9), and Schuftan (13) are of the opinion that the dilution process is generally superior.

Acknowledgment

The writer wishes to acknowledge the advice and comment of B. F. Dodge and C. A. Walker.

Notation

a = fraction of the extra 1 atm. air which is fed to the upper column proper; also relative volatility

ΔH = enthalpy of vaporization; also enthalpy change of separation, usually C.h.u./lb. mole

ΔS = entropy change of separation usually C.h.u./lb. mole, °K.

f = fraction of air fed to double column which may be at 1 atm.

H = enthalpy, usually C.h.u./lb. mole

M = number of moles

O = number of moles of overflow in a column

Q = heat effect in general

Q_{CB} = heat transferred in condenser-boiler

Q'', Q_T = special heat quantities defined in text

R = gas constant

T = temperature, usually °K.

T_o = ambient temperature, 300° K.

V = number of moles of vapor in a column

Fig. 4.
Process 6-III-A.

W = work

x = composition, usually mole fraction nitrogen of a liquid stream

y = composition, usually mole fraction nitrogen of a vapor stream

SUBSCRIPTS FOR $M, H, x,$ AND y

$_1$ = entering air

$_2$ = enriched air product

$_3$ = nitrogen waste

$_4$ = extra air stream

$_B$ = stream to boiler or to bottom column

$_C$ = overflow from column

$_F$ = feed to top column

$_L$ = liquid

$_O$ = overflow

$_R$ = reflux to top column

$_V$ = vapor

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Discussion

W. E. Lobo (Chairman) (The M. W. Kellogg Co., New York, N. Y.): Do you believe the analysis of yours shows the entire story in any case? If you are going to use 30% O_2 in an enriched-air mixture of 40 or 50%, you are obviously going to use it at some pressure, and therefore you would have to compress the products made in your process here or you would have to compress the product of the mixture of air and pure oxygen. I think when you look into that further, you will find that the handling of the minimum quantity of air—which is going to be, I believe, the air quantity with the high-purity oxygen—will show up still better, as oxygen can be withdrawn from a liquid-air fractionation unit under some pressure.

Harding Bliss: That is perfectly possible. I have not done it and cannot speak on it.

Charles F. Bonilla (Columbia University, New York, N. Y.): Your figures for energy requirement in horsepower hours per pound of total contained oxygen are, of course, in the standard units for cost estimates on alternative sources of oxygen. However, there might be some theoretical or other reason to be interested in the horsepower hours per pound of excess oxygen above and beyond the composition of air, which is, after all, available free. Figuring mentally, I note that in these latter units the dilution process would show to yet greater advantage compared with the direct process, for the compositions of each that you have given.

(Presented at Forty-first Annual Meeting, New York, N. Y.)

ELLIOTT OXGYEN PROCESS AND IMPURITY REMOVAL SYSTEM

D. B. CRAWFORD*

Elliott Company, Jeannette, Pennsylvania

MOST processes which have been proposed for the large scale production of "tonnage" oxygen are based on the Linde-Fränkl process which uses a double column for fractionation and reversing regenerators for purification of air (7). Modifications of the basic Linde-Fränkl process include the elimination of the high pressure (200 atm.) stream (1) (9), use of multifluid exchangers in place of regenerators (9), and use of an intermediate pressure column to improve oxygen recovery (1).

The Elliott oxygen process is designed to overcome certain difficulties associated with these processes; a technical description of the Elliott cycle and the operation of the various components follows.

Flow Sheet

A simplified flow sheet is shown in Figure 1. Flow quantities shown are moles of the various streams per mole of 95% oxygen product. In calculating the expander flow, heat leak was assumed to be 32 B.t.u./lb. mole of product, a figure based on pilot plant data, which would be applicable to a very large plant. Air is fed to the process by an air blower and is dried to a dew point of about −40° F. by means of a refrigeration unit and adsorption driers. The air then enters the clean-up exchanger system for final removal of water vapor, carbon dioxide, acetylene, and other condensable impurities. After passing through an accumulator, the air enters the column as a saturated vapor at a pressure of about 3 lb./sq.in. gage. (A section of this paper is devoted to a description of the impurity removal system.)

Plant Refrigeration

The fluid which is used for refrigerating the plant is nitrogen from the expander discharge and from the top of the column. This low pressure nitrogen

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is heated as two streams; one in the reflux cooler, exchanger G, and exchanger B, and the other in the clean-up exchangers. Nitrogen from exchanger B and nitrogen from the clean-up exchangers is compressed in a two-stage compressor to about 100 lb./sq.in. abs. The compressed nitrogen is cooled in exchangers A, B, and G. Between exchangers B and G the high pressure nitrogen is divided into a refrigeration stream to be fed to the expanders and a reflux stream to be sent through exchanger G and to the column reboiler where it is condensed. The refrigeration stream leaves the expanders as saturated vapor at −318° F. and joins the nitrogen stream leaving the top of the column.

Reflux nitrogen is liquefied in the column reboiler and is subcooled in the reflux cooler. Upon throttling to column pressure, about 7% of the liquid flashes to vapor. The oxygen stream, of 95% purity, is withdrawn as a saturated vapor from the bottom of the column and is warmed in exchanger A to a temperature of about 85° F.

In order to cool the air to its dew point before it enters the column, the incoming air is cooled entirely with nitrogen, while the oxygen product is used to cool part of the incoming compressed nitrogen. The clean-up cycle is so arranged that the nitrogen cycled through the clean-up exchangers is always kept free of condensable impurities. With this arrangement, together with the use of oil-free compressors, the compressed nitrogen maintains an impurity dew point of −318° F. In this way, deposition of impurities in exchangers A and B, expander nozzles, exchanger G and the reboiler is completely eliminated; and thus, long-term operation of the plant is assured.

This system produces all the oxygen and most of the nitrogen continuously as clean, dry products. A small portion of the nitrogen is used for clean-up purposes and carries out the impurities

brought in with the air. A small portion of the clean, dry compressed nitrogen is used for the operation of valves and instruments, for the provision of an inert atmosphere in the cold box, and for sealing at the shaft of the expander.

Impurity Removal System

The impurity removal system is designed to remove foreign particles and condensable impurities from the air feed to the plant and to discharge these impurities from the system in a manner that permits the plant to run indefinitely. Dust, cinders, and smoke are removed by a water wash.

The water-vapor content of the air is usually much higher than the carbon-dioxide content, and, in order to equalize the problem of removal in the clean-up exchangers, the bulk of the water vapor is removed with a cold water scrubber and subsequent adsorption on alumina. The final removal of water and carbon dioxide occurs in the clean-up exchangers where the air is cooled to −310° F. At this temperature, the vapor pressure of acetylene is so low that the quantity of acetylene which passes through the exchanger is not sufficient to constitute an explosion hazard. This small quantity of acetylene passing through the clean-up exchangers is reduced to an even lower value by adsorption on the silica gel in the accumulator.
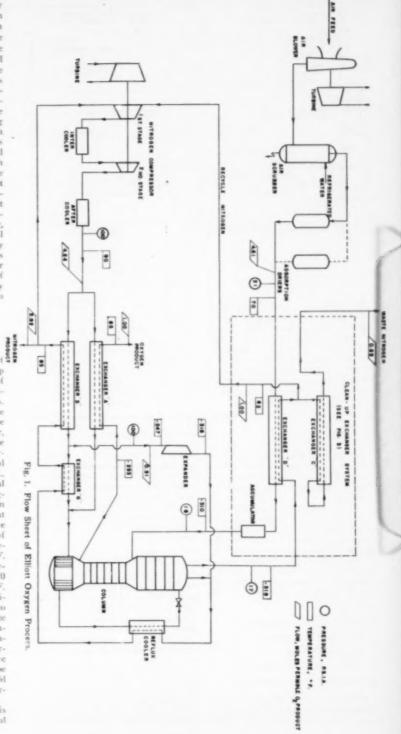
Steps in Clean-up Procedure

Fundamentally the clean-up exchanger system consists of operation of two gas-to-gas heat exchangers alternately. First, the air is fed to one exchanger until the deposition of impurities raises the pressure drop to an objectionable value. Then the plugged exchanger is switched out of operation while the idle exchanger is put into operation. The plugged exchanger is then derimed and made ready to replace the other exchanger as soon as it plugs.

The deriming is accomplished by passing clean, dry, waste nitrogen through the plugged exchanger: first, in the clean nitrogen passage, from the warm end to the cold end; then in the plugged air passage, from the cold end to the warm end, where it sublimes the deposits (*See* Fig. 2, plate 6). This flow path results in heat transfer between the incoming and outgoing nitrogen in the plugged exchanger, i.e., the clean dry nitrogen from the operating exchanger is cooled as it moves from the warm end to the cold end, and it is warmed as it returns from the cold end to the warm end. In comparison with common deriming procedures where the exchanger is warmed sufficiently to melt the deposits, this heat exchange minimizes the temperature rise of the heat exchanger mass. Thus, the work of refrigeration required, after the deriming, for cooling the exchangers to normal operating temperature becomes very small. In fact, designs of typical plants and pilot plant data show that the power required by the refrigeration losses of the clean-up operation represent only two per cent of the total power input to the plant.

Reducing Clean-up Nitrogen

In order to reduce the quantity of nitrogen which is required for cleanup of the fouled exchanger, the rate of sublimation of the impurities is increased by warming the exchangers. This is accomplished by allowing a time interval between the switching of the nitrogen flow to the derimed exchanger, and the switching of the air flow to the derimed exchanger. During this interval, the air stream warms the plugged exchanger uniformly about $36°$ F., resulting in the following typical figures which indicate the order of magnitude of the increase in sublimation rates. The vapor pressure of water at $0°$ F. is eight times its vapor pressure at $-36°$ F., and the vapor pressure of carbon dioxide at $-184°$ F. is twenty-two times its vapor pressure at $-220°$ F. Similarly the vapor pressure for acetylene at $-264°$ F. is approximately 370 times its vapor pressure at $-300°$ F. Thus the problem of removing impurities from the exchanger is reduced to the point where only 20% of the waste nitrogen is needed for cleanup. Advantage may be taken of this fact to economize on heat transfer surface by arranging, for example, a bank of five exchangers, of which four would be operating and one cleaning. This would result in a ratio of nonoperating surface to operating surface of only 0.25.

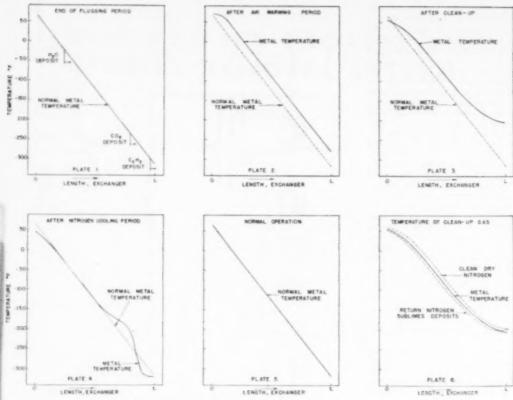The derimed exchanger, before it is put into operation, is cooled to normal

Fig. 1. Flow Sheet of Elliott Oxygen Process.

END OF PLUGGING PERIOD

H_2O DEPOSIT

NORMAL METAL TEMPERATURE

CO_2 DEPOSIT

C_2H_2 DEPOSIT

PLATE 1

LENGTH, EXCHANGER

AFTER AIR WARMING PERIOD

METAL TEMPERATURE

NORMAL METAL TEMPERATURE

PLATE 2

LENGTH, EXCHANGER

AFTER CLEAN-UP

METAL TEMPERATURE

NORMAL METAL TEMPERATURE

PLATE 3

LENGTH, EXCHANGER

AFTER NITROGEN COOLING PERIOD

NORMAL METAL TEMPERATURE

METAL TEMPERATURE

PLATE 4

LENGTH, EXCHANGER

NORMAL OPERATION

NORMAL METAL TEMPERATURE

PLATE 5

LENGTH, EXCHANGER

TEMPERATURE OF CLEAN-UP GAS

CLEAN DRY NITROGEN

METAL TEMPERATURE

RETURN NITROGEN SUBLIMES DEPOSITS

PLATE 6

LENGTH, EXCHANGER

Fig. 2. Schematic Temperature Behavior of Clean-up System.

operating condition by means of the cold nitrogen, which is passed through the derimed exchanger simultaneously with the air warming of the plugged exchanger.

As an example of the magnitude of the time intervals involved, the pilot plant exchangers use a two-minute air warming period and a normal operating period of four hours.

For a summary of the flow patterns involved in one complete cycle for either exchanger, reference is made to Figure 3. The cycle consists of the following steps:

Plate 1—Normal, plugging
Plate 2—Air Warming
Plate 3—Cleanup
Plate 4—Nitrogen cooling
Plate 1—Normal, plugging

Calculation of the temperature distribution curves of the exchangers during each step of the impurity removal period is a complex problem in unsteady state heat transfer and conduction, which does not lend itself to a simple analytical solution. The problem was solved graphically, with a modified Schmidt method, and the temperature distribution after each step in the operation may be seen in Figure 2, plates 1 to 5.

Impurity Accumulation

At an air temperature of $-310°$ F., the vapor pressure of carbon dioxide is so low that only 10^{-9} volumes of carbon dioxide will be carried through the exchanger with one volume of air. Indeed, in a 100-ton/day oxygen plant, this represents an accumulation of less than 1 lb. of carbon dioxide in 365 days of operation.

Even though the clean-up exchangers cool the air to $-310°$ F. before it enters the column, not all the acetylene is deposited in the exchangers. For example, based upon an extrapolation of data of Burho (2), vapor pressure of acetylene

at $-310°$ F. is 1.35×10^{-5} mm. Hg. This corresponds to 1.4×10^{-8} volumes of acetylene per volume of air leaving the clean-up exchangers. During each warm-up period, the average concentration of acetylene for the two-minute period rises to 150×10^{-8} mole of acetylene per mole of air leaving the clean-up exchangers. The warm-up period occupies two minutes out of four hours. Assuming, for the moment, that no acetylene is removed in the accumulator and that the volume of air entering the column is five times the volume of oxygen produced, the acetylene concentration in the oxygen leaving the column would have to be $5 \times (1.4 + 1.25) \times 10^{-8} = 13.2 \times 10^{-8}$ moles of acetylene per mole of oxygen, in order to carry out all the acetylene that the air brings in.

At $-293°$ F. the boiling temperature of the oxygen in the column, the vapor pressure of acetylene is 3.2×10^{-4} mm.
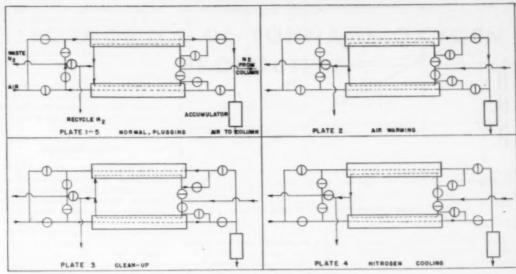
Fig. 3. Clean-up Cycle Flow Sheet.

Hg. which corresponds to 32.4×10^{-8} moles acetylene per mole of oxygen leaving the reboiler. If the amount of acetylene being brought into the column exceeded this quantity, then the acetylene would tend to accumulate in the liquid oxygen as a solid, and would represent a potential hazard. Actually, only $(13.2/32.4 = 0.41)$ 41% of the dangerous quantity of acetylene will leave the exchanger. Figure 4 shows the relation of the per cent of dangerous concentration and the air temperature leaving the clean-up exchangers. In actual operation, the silica gel in the accumulator, which is placed between the exchangers and the column, adsorbs both the acetylene and carbon dioxide to such an extent that not even a trace of either can be found in the oxygen product.

Cycle Characteristics

The separate nitrogen circuit makes it possible to control the reflux rate independently of the air rate. Thus the optimum quantity of reflux may be supplied to meet the variable demand corresponding to different plant capacities or different product purities.

To show the characteristics of the single column, calculations of the recovery as a function of reflux were made for a column of 18 theoretical plates producing 95% gaseous oxygen. The average pressure in the column was assumed to be 18 lb./sq.in. abs. Composi-

tion of air was assumed to be: 20.95% O_2; 78.12% N_2; 0.93% A. The equilibrium data for the oxygen-nitrogen-argon system, required for calculating the columns, were obtained from a recent correlation (3) based primarily on the work of Dodge and Dunbar (6). Enthalpy data were obtained from the recent charts of Claitor and Crawford (4).

The curve giving the moles of the oxygen produced per mole of liquid reflux vs. per cent recovery is shown in Figure 5. Also shown in Figure 5 is the oxygen content of the nitrogen leaving the top of the column for various recoveries. It is apparent that the yield of oxygen per mole of reflux does not fall off rapidly until the oxygen recovery is considerably more than 98%. Economic balances based on the production of 95% O_2 alone indicate that the least cost per unit of oxygen is obtained when the nitrogen produced as the by-product contains less than 0.5% O_2. This is a different characteristic from that exhibited by the Linde-Fränkl process, where operation of the reversing regenerators with the double column usually results in a nitrogen product stream containing 2% O.

Typical temperature differences in the heat exchangers using the warm-end temperatures of Figure 1 are plotted in Figure 6. The temperature difference in the clean-up exchanger is practically constant at 8° F. because the pressure of both streams in the clean-up exchanger

is close to atmospheric, and the difference in Joule-Thomson effect is negligible. The temperature difference in the oxygen exchanger varies primarily because of the larger Joule-Thomson effect in the high pressure nitrogen. The temperature difference in exchanger B
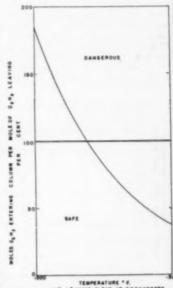

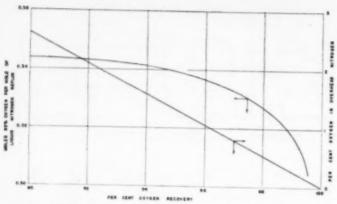Fig. 4. Allowable Acetylene Entering Column.

Fig. 5. Oxygen Recovery for 18 Theoretical Plates.

is almost identical with the oxygen exchanger, *A*, down to −247° F. at which temperature the nitrogen is fed to the expander.

Since the high pressure exchangers are kept perfectly free of impurities, the large temperature differences at the cold end of the exchangers can be fully utilized in terms of saving in heat exchanger surface.

Automatic control of the plant is achieved in part by separate control of the air feed and the nitrogen compression. Division of the nitrogen stream is controlled further so that the correct proportion of the compressed nitrogen goes to the expander to maintain the proper refrigeration balance, and the exact quantity of nitrogen goes to the column for most economic recovery. With this arrangement, the plant can be operated at reduced capacity without loss of efficiency. Switching operations of the clean-up exchangers are con-

trolled by automatic timers on a forced cycle.

This plant is suited not only for the production of 95% gaseous oxygen, but also for oxygen of 99.5% purity and for liquid oxygen. In these applications, the clean-up system operates in an identical manner, and the product recovery is likewise high. In special applications, where the over-all height of the plant must be kept to a minimum, the single column offers an obvious advantage over the double column.

Since the original conception of this cycle, a pilot plant has been erected and operated at the Jeannette (Pa.) works of the Elliott Co. (*12*). Operation was entirely successful and confirmed the advantages of the cycle. Two runs were made in which the pilot plant operated continuously for more than two months. Inspection of the component parts indicated no reason why the plant could not run indefinitely. Acetylene was found

in the air entering the plant but no trace of acetylene was detected in frequent tests on the liquid oxygen in the reboiler. The product and 80% of the waste nitrogen were produced continuously as clean dry streams (dew point approx. −300° F.).

Applications for patents on the impurity removal system and on the entire oxygen cycle have been filed by the author, to be assigned to the Elliott Co.

Acknowledgment

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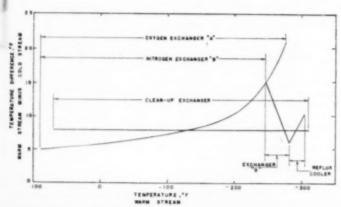
Fig. 6. Heat Exchanger for Temperature Differences.

ECONOMICS OF TONNAGE OXYGEN PRODUCTION

IRVING ROBERTS

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AS a result of the recognized advantages of the use of tonnage oxygen in the chemical, metallurgical, gas and petroleum industries (6), eight tonnage plants are in the process of erection or are already in operation in the United States. Capacities of these plants range from 120 tons/day to 1000 tons/day. Uses for the oxygen include the manufacture of steel, oxygenated chemicals, synthetic gasoline, and synthetic ammonia.

So far as is known, all these plants are based upon the Linde-Fränkl process (7) or its modifications (1, 9). The Elliott process is designed to overcome certain disadvantages of the Linde-Fränkl process, particularly with regard to continuity of operation and to acetylene explosion hazard. Other papers on the Elliott process describe the principles of the cycle (5), construction and operation of the pilot plant (10), and the gas-to-gas heat exchangers (12). The present paper is devoted to a study of the design of a tonnage plant, and to the economics of oxygen production by the Elliott process.

The plant chosen for this design study produces 120 tons/day of 95% oxygen gas. The plant is assumed to be located in the Pittsburgh area, using steam for the main drives, and electric power for auxiliaries.

A flow sheet of the cycle is given in Figure 1. The cycle consists of the major features which characterize the Elliott process, i.e., an air purification system in which the air is cooled at low pressure to approximately its dew point before it enters a single column, and a nitrogen circuit which supplies high pressure nitrogen in two streams, one of which flows to an expander for refrigeration, and the other to the column for reflux. However, comparison of Figure 1 with the flow sheet of the pilot plant (10) will show a major change in the clean-up heat-exchanger system.

The new arrangement takes advantage of the fact that a relatively short period is required for cleanup. In the pilot plant, two heat-exchanger units were provided; one was cleaned while

the other was fouled during the cooling of air. For a fouling period of 4 hr., it was found that the exchanger was completely cleaned in 45 min., when the entire by-product nitrogen stream was used for cleaning. Thus, each heat exchanger was idle for a period of 3¼ hr. during successive cycles. In an alternative method of operation, only 19% of the by-product nitrogen stream was used for cleanup, and the cleaning was continued for the entire 4-hr. period. This method has the advantage of making available continuously 81% of the nitrogen by-product as a clean, dry stream, as well as reducing the pressure drop in the clean-up nitrogen circuit.

In the present plant (Fig. 1) four banks of heat exchanger units, C, D, E and F, are provided for the clean-up system. Of these, three are operating in parallel in air cooling service, while the fourth is being cleaned. A switching occurs once every hour, so that each heat exchanger is fouled for a period of 3 hr. and then is cleaned for a period of 1 hr. In Figure 1, as an example, the arrows showing direction of flow indicate that exchanger C is being cleaned, while exchangers D, E and F are being fouled. Provision of four exchanger banks in this manner results in a saving in heat-transfer surface, since the ratio of non-operating surface to operating surface is reduced from 100% in the pilot plant to 33⅓% in this plant.
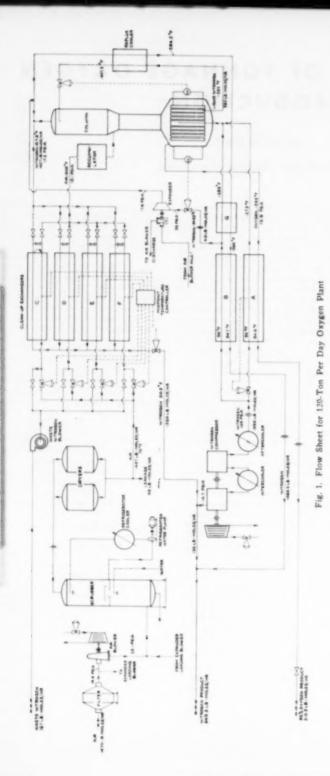
As a consequence of this arrangement, the exchangers which are operating in parallel will be fouled to different extents. For example, exchanger D might have just been switched into operation after cleaning, while exchanger E might have been fouled for a 1-hr. period, and exchanger F might have been fouled for a 2-hr. period. Therefore the three exchangers will contain different quantities of impurity deposits and will offer different resistances to air flow. To assure balanced flow in all three operating exchangers, automatic control valves are placed in the individual air lines, and these are controlled from the midpoint temperatures of the heat exchangers. The midpoint temper-

ature is used as the control variable since it is sensitive to flow unbalance; i.e., a 1% change in flow balance results in approximately a 20° F. change in midpoint temperature.

In Figure 1, a waste nitrogen exhaust blower is used to draw the clean-up nitrogen from the clean nitrogen manifold at the warm end of the clean-up exchangers, through the exchanger being cleaned. The alternative to the use of an exhaust blower would be to provide additional head in the air blower to overcome the resistance to flow in the clean-up circuit, so that the waste nitrogen may be discharged at atmospheric pressure. With the exhaust blower, this additional head is applied only to the clean-up nitrogen flow, which is much smaller than the air flow, thus effecting a saving in power. It will be noted that about 15% of the by-product nitrogen is used for clean-up in this plant, compared with the 19% figure used in the pilot plant. This reduction is obtained from the more favorable ratio of specific volumes of the air and nitrogen, which results from the somewhat larger difference in pressure between the two streams in the present clean-up system.

Economic Factors in Design

In a tonnage oxygen plant, the major item of operating cost is power for compression. Additional items, such as cooling water and operating labor, have little influence on problems involving choice of low temperature process conditions, and are neglected in the analysis given below. Therefore, the economic phases of the design are resolved essentially into a balance of fixed charges against power costs.

For this analysis, power is valued at 0.4 cents/hp.hr., roughly equivalent to electric power at 0.5 cents/kw.hr., and to steam, used in condensing turbines, at 40 cents/M lb. Fixed charges are taken at 25% of investment cost yearly, and this figure is intended to include depreciation, taxes, insurance and maintenance. Maintenance is included in this figure since usually it is assumed to be

Fig. 1. Flow Sheet for 120-Ton Per Day Oxygen Plant

a fixed percentage (1.5 to 3%) of the investment cost yearly.

A factor which will be useful in the economic balance calculations is the value of low temperature refrigeration. A discussion of this factor is given in the next section of this paper, and this is followed by an economic analysis of the following process variables: optimum warm-end temperature differences, optimum nitrogen compressor discharge pressure, optimum number of column trays, and optimum oxygen recovery. These process variables are believed to be the most important for design of the plant. Many other economic balances are required for complete design, but these are either of minor importance or give well-known results. With the exception of the cost of refrigeration they are not considered in this paper.

Flows, temperatures and pressures shown in Figure 1 are the result of this economic study.

Cost of Refrigeration

Refrigeration is supplied by the turboexpander to overcome the effects of heat leak and warm-end temperature loss. Expanders are usually installed in duplicate for rapid cooling down of the plant, and one of these is cut off after the desired column liquid levels are reached. In a steadily operating plant, any deficiency or excess of refrigeration results in a decrease or increase in inventory of liquid in the plant. This appears as a change in the level of liquid oxygen in the reboiler, and, therefore, this level is made the basis of refrigeration control by automatic throttling of the nitrogen flow to the expander. In order to keep the throttling loss to the minimum required for control, the expander nozzle ring is sized to pass the required flow at the desired pressure drop. Since, in designing a plant, there is always considerable uncertainty in the heat leak figure, a set of nozzle rings of graded sizes is provided. The largest nozzle ring is used in starting the plant, and after steady operation has been achieved, the proper nozzle ring is calculated and installed.

The power output of the expander in normal operation is usually about 2 to 4% of the power input to the plant. This is absorbed by an air blower which operates in parallel with the main air blower, and thus reduces the load on the latter.

For use in later calculations, it is now desirable to calculate the cost of 1000 B.t.u. of refrigeration. In this analysis, it is assumed that a molar balance is maintained in each of exchangers A and B, and in the clean-up exchangers. Assuming an expander efficiency of 80% and an inlet pressure of 95 lb./sq.in.abs., the enthalpy drop, using the charts of

Claitor and Crawford (*4*), is 420 B.t.u./lb.mole. Not all this is useful refrigeration, however, since each additional pound mole of nitrogen passed through the expander entails an additional warm-end temperature loss at exchanger *B*. From the *T-S* diagram, this is 20 B.t.u./lb.mole, leaving 400 B.t.u. as the net useful refrigeration/pound mole through the expander. On this basis, each 1000 B.t.u. of refrigeration requires the compression of 2.50 lb. moles of nitrogen. Refrigeration in the additional cold nitrogen leaving the expander is imparted to the system by subcooling of liquid reflux in the reflux cooler, and by cooling of high pressure nitrogen in exchangers *G* and *B*.

Whether the 2.50 figure can be used to determine the cost of a particular item of heat leak or of warm-end temperature loss depends upon the position at which this refrigeration requirement occurs. For example, one might consider 1000 B.t.u. of heat leak occurring at the top tray of the column. This heat leak has the effect of vaporizing liquid nitrogen from the tray, tending to reduce the quantity of liquid reflux available for operation of the column. Therefore, to maintain the reflux flow at the desired rate, additional liquid nitrogen must be supplied to the top of the column, and this results in an additional nitrogen compression requirement beyond that required for the expander.

The additional nitrogen flow for reflux is calculated as follows: The latent heat of nitrogen at 17.2 lb./sq.in.abs. is 2380 B.t.u./lb.mole, so that the 1000 B.t.u. of heat leak vaporizes 1000/2380 = 0.42 lb.mole of liquid. Allowing for 6% flash at the reflux throttle valve, a value which is obtained with a reasonable amount of reflux cooler surface in a plant of this capacity, the additional quantity of liquid leaving the reflux cooler becomes 0.44 lb.mole. This 0.44 lb.mole also increases the warm-end temperature loss at exchanger *B*, resulting in 0.44 × 20 × 2.50/1000 = 0.02 lb. moles of further nitrogen to be compressed. Thus, the presence of 1000 B.t.u. of heat leak at the top of the column requires the compression of 2.50 + 0.44 + 0.02 = 2.96 lb. moles of nitrogen.

For heat leak at the bottom of the column, no additional reflux is required. In fact, there is an increase in reflux entering the top tray of the column due to more effective subcooling of reflux by the increased cold nitrogen flow from the expander, which reduces the percentage of flash upon throttling. However, this is a slight effect which is neglected (the temperature to which the reflux is subcooled is assumed constant) and the 2.50 figure is assumed to apply to 1000 B.t.u. of heat leak at the bottom of the

column. This same figure holds for heat leak into exchangers *A*, *B* and *G*, heat leak into the expanders, and warm-end temperature loss at exchangers *A* and *B*.

For warm-end temperature loss and heat leak in the clean-up exchanger system and the accumulator, the effect is equivalent to heat leak at the feed tray of the column. In this situation, feed tray liquid is evaporated, so that the liquid reflux has been effective in the upper half of the column before it is evaporated. Evidently, 1000 B.t.u. of refrigeration for this purpose will require the compression of some quantity between 2.50 and 2.96 lb.moles of nitro-

gen. As a reasonable estimate, it will be assumed that 2.73 lb.moles, the mean of the above two figures, applies to 1000 B.t.u. of refrigeration requirement in the clean-up exchanger system.

In the economic balances given in the body of this paper, interest is primarily in the value of warm-end temperature loss in the clean-up exchanger system on the one hand, and in exchanger *B*, on the other. Therefore, the costs of both types of refrigeration are calculated, for use at the appropriate points.

In either type of refrigeration, the additional nitrogen flow requires an in-

crease in capacity of the compressor, the expander, exchanger B, exchanger G, and the reflux cooler. The last two items are relatively small exchangers, and their increase in first cost is neglected. For the nitrogen compressor with all accessories the installed cost is taken at $60/hp., and for the expander, at $150/hp. Heat exchanger B is assumed to consist of standard sections (to be discussed below), loaded at the rate of 60 lb. moles/(hour)(section), and valued at $3200/section, including manifolds, connections and installation. The adiabatic stage efficiency of the nitrogen compressor is taken at an average of 67%, based upon typical American multistage centrifugal compressors in this size range. The power output of the expander is assumed to be transmitted to the loading blower with an efficiency of 85%, to allow for transmission and bearing losses. The investment cost of the main air blower is unchanged, since it is assumed that this is sized to take the full air load, independent of the portion normally taken by the expander loading blower. It is assumed that the plant operates 360 days (8640 hr.)/year.

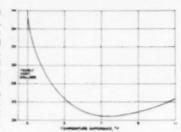Considering first the case of warm-end loss in the clean-up exchangers, the

Fig. 2. Optimum Temperature Difference for Clean-Up Exchanger. Preliminary analysis.

nitrogen compression requirement is 2.73 lb.moles/hr. at a refrigeration rate of 1000 B.t.u./hr. For this the investment costs are:

Nitrogen compressor: 3.85 hp. at $\,$ 66 = $254
Expander   0.414 hp. at $\,$ 150 =   62
Expander B: 0.0455 section at $3200 = 146
                                      ————
                    Total investment = $462

The yearly cost is given by:

Power: (3.85 -- 0.35)
            hp.hr. × 8640 × $0.004 = $122
Fixed charges: $462 × 0.25 = 116
                                      ————
   Total yearly cost of 1000 B.t.u./hr. = $238

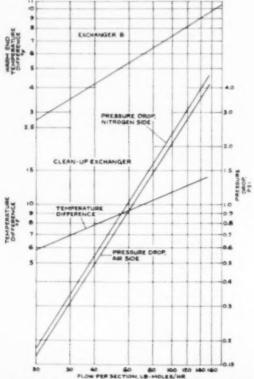This is equivalent to a cost of 2.76 cents/1000 B.t.u. of refrigeration.

Fig. 3. Performance of Standard Heat-Exchanger Sections.

For the case of warm-end loss in exchanger B, for which 2.50 lb.moles/hr. of nitrogen are compressed, a similar calculation gives a yearly cost of $216 for 1000 B.t.u./hr. of refrigeration, based upon an exchanger B loading rate of 60 lb.moles/(hr.)(section). For other exchanger B loading rates, refrigeration costs are calculated in the first part of Table 3.

Optimum Warm-end Temperature Difference

This problem will be considered first for the case of the clean-up exchangers which require the majority of the gas-to-gas heat-transfer surface in the plant. In an exchanger operating in this service, the air and nitrogen streams are both at low pressure, so that their specific heats are almost identical at the same temperature. Thus, for balanced flow, the temperature difference between the two streams is essentially constant throughout the entire length.

In the early part of this work, a rough approach to this economic balance was made by neglecting all factors except the cost of refrigeration and the fixed charges on the heat-transfer surface. For an analysis of this type, it is assumed that gas-to-gas heat exchangers are available with extended surface in both the air and nitrogen passages. The over-all coefficient of heat transfer per unit of direct surface is estimated at 60 B.t.u./(hr.)(sq.ft.)(° F.), and the installed cost is estimated at $6.00/sq.ft. of direct surface. On this basis, results given in Table 1 and plotted in Figure 2, indicate an optimum temperature difference in the range of 7 to 8° F.

It was on the basis of a crude analysis of this type that preliminary requirements were established for the development of mass-production heat-exchanger units (12) for use in all sizes of tonnage plants. As finally developed, these standard units are 10 ft. long, and two of these units in series are considered as a standard section. Any number of such standard sections may be manifolded in parallel for use either in clean-up exchanger service, exchanger B service, or exchanger A service. The heat-transfer and pressure-drop performance of a standard section in clean-up exchanger service is shown in the lower part of Figure 3.

As these standard heat-exchanger sections have been adopted, the problem now becomes one of determining the optimum loading rate of these sections in clean-up exchanger service. An increase in loading rate, which reduces the number of sections required, increases not only the refrigeration required to overcome the warm-end temperature loss, but also the air blower

power required to overcome the frictional resistance to air and nitrogen flow in the exchangers. Table 2 shows an economic balance of these factors, including cost of air blower, cost of air-blower power, cost of refrigeration to overcome warm-end and clean-up loss, and cost of exchangers. Minor factors which are neglected in this table are the changes in the cost and power requirement of the nitrogen exhaust blower, and changes in cost of the air-scrubbing and dehumidification system due to the change in air-blower pressure. The change in flow resistance also changes the pressure in the column, thus changing the temperature difference in the reboiler. The resulting change in the cost of the reboiler surface has been found to affect this analysis only slightly, and is therefore neglected for simplicity.

Results are plotted as the upper curve of Figure 4, and indicate an optimum loading rate of about 68 lb.moles/(hr.) (section) corresponding to a temperature difference (Fig. 3) of 9.8° F.

The loading rate of exchanger B is restricted by the pressure drop in its low pressure side. This may be seen from Figure 1, in which nitrogen from the top of the column is returned to the inlet of the nitrogen compressor by two paths. The first path consists of flow through the clean-up exchangers, and the second consists of flow through the reflux cooler, through exchanger G, and through exchanger B. After the loading rate has been established, and therefore the pressure drop on the nitrogen side, in the clean-up exchangers, the maximum loading rate in exchanger B is then determined as that which will just make the pressure drops of the two paths the same. Using designs of the reflux cooler and of exchanger G which require only slight pressure drops on the low pressure side, the maximum loading rate of exchanger B is found to be 60 lb.moles/(hr.)(section).

However, an economic analysis of the exchanger B loading rate must still be made, since it is possible that the optimum rate will be lower than 60 lb.moles/(hr.)(section). This is carried out in Table 3, on the basis of the warm-end temperature difference curve given in the upper part of Figure 3. Pressure drop on the high pressure side of exchanger B is a negligible factor in these calculations.

In Table 3, the change in warm-end temperature loss due to change in loading rate, results in a change in the net refrigeration per pound mole of nitrogen flow through the expander. Therefore, the first part of this table is devoted to a recalculation of the cost of refrigeration for the various loading rates by the method given earlier in this
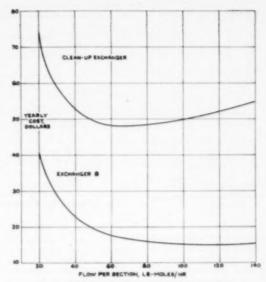

Fig. 4. Optimum Heat-Exchanger Loading Rates.

paper. The second part of Table 3 determines the yearly costs of flow of 1.0 lb.mole/hr. through exchanger B as the sum of the yearly cost of the exchangers and the yearly cost of refrigeration required to overcome warm-end temperature loss at B. Results are plotted in Figure 4, and indicate an optimum flow rate considerably in excess of 60 lb. moles/(hr.) (section). Hence, the loading rate of exchanger B is set at 60 lb. moles/(hr.) (section) on the basis of the pressure drop limitation.

Because of the pressure drop interrelationship between exchanger B and the clean-up exchangers, a more rigid analysis would determine the optimum loading rates of both units simultaneously. In this method, the loading rate of exchanger B would be adjusted to correspond to the assumed loading rates of the clean-up exchangers, and after weighting by the appropriate flows, total costs would be determined. Additional refinements would correct the cost of the reboiler for changing column pressure, and the value of refrigeration for changing back pressure on the expander. However, all these factors are relatively small and apparently cancel each other, so that the more accurate analysis yields about the same optimum loading rates as found here.

For the case of exchanger A, pressure drop on the low pressure (oxygen) side will again be an important factor. For most processes in which oxygen is used, the oxygen is compressed to a higher pressure. In such cases, pressure drop in exchanger A will have value in

terms of compression power, and this can only be evaluated in terms of the particular process involved. In view of this uncertainty, and because the surface requirement of exchanger A is relatively small, the simplifying assumption is made that the loading rate of exchanger A is approximately the same as that of exchanger B. Since the performance of a standard section in exchanger A service is close to its performance in exchanger B service, the assumption of equal loading rates implies that the warm-end temperature differences will also be approximately the same.

Optimum Compressor Discharge Pressure

The economic balance involved in the determination of the optimum compressor discharge pressure may be visualized as follows: An increase in the compressor discharge pressure increases the expander pressure ratio, and therefore increases the enthalpy drop per pound mole (expander efficiency assumed constant at 80%). Simultaneously, the higher pressure causes a decrease in warm-end temperature difference at exchangers A and B, resulting in a smaller warm-end temperature loss. Thus, the expander nitrogen flow is decreased for the same plant capacity, decreasing the surface requirement of exchanger B, and decreasing the expander power. However, the nitrogen flow required for reflux is unchanged, and the over-all decrease in nitrogen
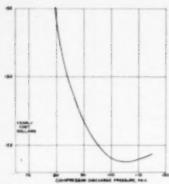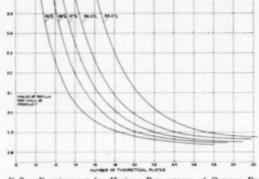
Fig. 5. Optimum Compressor Discharge Pressure.

Fig. 6. Reflux Requirement for Various Percentages of Oxygen Recovery.

flow does not compensate for the increase in pressure ratio at the compressor, i.e., there is a net increase in compressor cost and power consumption.

The other large effect occurs at the reboiler of the fractionating column. The increase in nitrogen pressure causes an increase in the reboiler temperature difference. This results in a very marked decrease in surface requirement, particularly because the heat transfer coefficient in the reboiler rises as an exponential function of the temperature difference. There are other factors which have a minor influence in this problem, namely, the changes in cost of the reflux cooler and of exchanger G, and changes in heat leak. These factors are again neglected.

Calculations are made on the basis of 1.0 lb.mole/hr. of nitrogen flow to the reboiler, and are summarized in Table 4. Results are plotted in Figure 5, and indicate an optimum compressor discharge pressure at about 105 lb./sq.in.abs.

It should be noted that the compressor costs per horsepower and the compressor efficiencies given in Table 4 apply to compressors suitable for a 120-ton oxygen plant, i.e., capacity of about 10,000 inlet cu.ft./min. As the capacities of such machines are increased, their cost per horsepower decreases, and their efficiency increases. Thus, for a compressor of 80,000 inlet cubic feet per minute, the size suitable for a 1000-ton plant, the installed cost is estimated at about $25 per horse-

power, and the average stage efficiency at about 73%. In the calculations given in Table 4, the effects of decreased compressor cost and increased compressor efficiency tend to increase the optimum compressor discharge pressure slightly as the plant capacity is increased.

Optimum Number of Column Plates

For solution of this problem, an extensive series of calculations was carried out to determine the reflux requirement as a function of number of trays for the single column. The oxygen purity was held constant at 95.0 mole %, and the oxygen recovery, which is defined as the moles of oxygen appearing in the product/100 moles of oxygen in the air feed, was varied from 90 to 99.5%. Relative volatilities were based on a recent correlation of Claitor and Cost (5).

Since the presence of argon cannot be neglected in these calculations without appreciable error, the three-component system was handled by a tray-by-tray method, using fictitious molecular weights to compensate for differences in latent heat. This method was found to be in excellent agreement with several rigorous tray-by-tray calculations, in which heat and material balances were carried out on each tray.

Results are plotted in Figure 6, in which the ordinate represents the moles of liquid reflux entering the top tray per mole of total 95% oxygen product.

For each of the curves of Figure 6, the optimum number of trays is determined as an economic balance of the cost of the trays against the cost of supplying liquid reflux to the column. Such calculations were carried out for each of the oxygen recoveries, and an example is given in Table 5 for the case of 90% recovery. In this table, the cost of the trays is taken to be proportional

TABLE 4. OPTIMUM COMPRESSOR DISCHARGE PRESSURE

Costs per lb mole per hour of nitrogen flow to reboiler

Compressor discharge pressure, lb./sq.in.abs.	71	82	93	104	115
Expander inlet pressure, lb./sq.in.abs.	65	75	85	95	105
Expander enthalpy drop, B.t.u./lb.mole	318	356	389	420	451
Warm end temperature difference, °F.	7.3	6.7	6.0	5.3	4.6
Warm end loss, B.t.u./lb.mole	41.1	34.8	27.8	20.2	11.8
Net expander refrigeration, B.t.u./lb.mole	277	321	361	400	439
Refrigeration required, lb.mole of reboiler nitrogen, B.t.u.²	183	177	170	162	154
Nitrogen compressed, lb.mole of reboiler nitrogen, lb.moles	1.661	1.551	1.471	1.405	1.351
Exchanger sections required²	0.0277	0.0258	0.0245	0.0234	0.0225
Per cent of nitrogen flow through expander	39.8	35.5	32.0	28.8	26.0
Compressor average stage efficiency, %³	67.8	67.4	67.3	67.2	67.0
Compressor power, lb.mole of reboiler nitrogen, hp.	1.82	1.88	1.93	1.98	2.03
Expander power, hp.⁴	0.082	0.078	0.072	0.067	0.062
Net power consumption, hp.⁵	1.78	1.81	1.87	1.92	1.98
Installed cost of compressor, $/hp.⁶	82.60	63.90	64.90	65.80	66.70
Reboiler nitrogen pressure, lb./sq.in.abs.	69	80	90	101	112
Reboiler temperature difference, °F	3.2	4.6	7.4	10.4	12.9
Reboiler heat transfer coefficient, B.t.u./hr.(sq.ft.)(°F.)⁷	50	74	97	157	202
Reboiler surface, sq.ft.	35.5	6.23	2.98	1.49	0.82
Installed cost of compressor, $	113.90	119.90	125.20	130.20	133.40
Installed cost of expander, $⁶	12.30	11.70	10.80	10.00	9.30
Installed cost of exchangers, $	88.70	82.80	78.40	74.90	72.00
Installed cost of reboiler, $⁷	355.00	62.50	29.60	14.90	8.20
Total investment cost, $	569.90	276.70	244.00	230.00	224.90
Fixed charges, $/yr.	142.48	69.18	61.00	57.50	56.23
Cost of power, $/yr.	60.50	62.60	64.60	66.30	68.10
Total yearly cost, $	202.98	131.78	125.60	123.80	124.33

¹ Exchanger B loaded at 60 lb.moles/hr.(section)
² Includes all plant heat leak and warm end loss, except warm end loss on expander flow. Warm end ΔT at exchanger A assumed identical with that at exchanger B.
³ Multistage centrifugal compressor, turbine-driven, one intercooler.
⁴ 85% of expander power credited.
⁵ Based on outside tube surface, estimated from pilot plant data.
⁶ $150 hp.
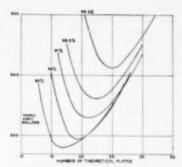⁷ $10/sq.ft. of outside tube surface.

Fig. 7. Optimum Number of Trays for Various Percentages of Oxygen Recovery.

to the flow of top tray vapor, i.e., a constant linear velocity is assumed. Also, it should be noted that the vapor rate will be considerably higher above the feed than below the feed, so that two different tray diameters would probably be used for each column. The cost per tray used in Table 5 is intended to be an average figure to cover both tray diameters. Fortunately, for all columns, the number of trays below the feed is approximately equal to the number of trays above the feed, so that the use of an average tray cost is justified.

The cost of supplying reflux to the column includes the cost of the reboiler, cost of the compressor and the cost of compression power for the reflux nitrogen. In addition, the cost of exchanger B to handle the reflux nitrogen, and of the refrigeration to overcome the warm-end loss at B, must be included. In Table 5, the latter two items are handled together, using the yearly cost from Table 3 for this purpose. As before, the costs of the reflux cooler and of exchanger G are neglected. Also neglected in these calculations are the effects of the changes in pressure drop in the column resulting from changes in the number of trays.

Results are plotted in Figure 7 from which it can be seen that the optimum number of theoretical plates increases from seventeen at 90% recovery to twenty-five at 99.5% recovery.

Optimum Oxygen Recovery

The optimum recovery is obtained by adding to the minimum costs from Figure 7, the costs of supplying the quantities of air corresponding to each oxygen recovery. The cost of supplying one pound mole of air may be determined as follows: The minimum yearly cost given by the upper curve of Figure 4 includes the cost of the air blower, the air-blower power, the clean-up exchangers, and the refrigeration required for the warm-end and clean-up loss at

the clean-up exchangers. To this is added the installed cost of the dehumidification system, which is estimated at $96/(lb.mole of air)(hr.), and the power required for its operation, which is estimated at 0.077 hp.hr./lb.mole of air. There is:

```
Yearly cost from Figure 7 ...........$48.60
Fixed charges, dehumidification system
  = $96 × 0.25 .....................  24.00
Yearly power cost, dehumidification sys-
  tem = 0.077  hp. × 8640 × 0.004     2.66
                                      ───────
Total yearly cost/(lb.mole of air)(hr.)...$75.26
```

This result has been used in the preparation of Table 6, and results are plotted in Figure 8. An optimum recovery of 98.5% is indicated.

The 120-Ton Plant. The cycle for the 120-ton plant (Fig. 1), has been calculated on the basis of results given above as follows:

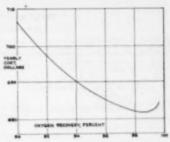|	
Production rate, 95% oxygen	= 310.5 lb.moles/hr.
Contained oxygen = 310.5 × 0.95	= 295 lb.moles/hr.
Air rate = 295/0.985 × 0.21	= 1427 lb.moles/hr.
Warm-end loss, clean-up exchangers = 9.8° F. × 6.96 × 1427	= 97,300 B.t.u./hr.
Clean-up loss = 97,300 × 0.18	= 17,500 B.t.u./hr.
Heat leak, estimated	= 30,000 B.t.u./hr.
Total losses affecting reflux rate	= 144,800 B.t.u./hr.

Fig. 8. Optimum Oxygen Recovery.

Based upon the column calculations, the composition of the products is the following:

	Oxygen Product %	Nitrogen Product %
Nitrogen	2.1	99.2
Argon	2.9	0.4
Oxygen	95.0	0.4

In the column calculations mentioned above, it was assumed that the air is fed to the column as a saturated vapor. Since the nitrogen leaves the top of the column as a saturated vapor, there is a temperature difference of about 10° F. in the clean-up exchangers for which no extra reflux need be supplied. This is equivalent to a warm-end loss of 10° F. × 6.96 × 1427 = 99,300 B.t.u./hr.

Total losses requiring extra reflux	= 45,500 B.t.u./hr.
Extra reflux required = 45,500/2380 × 2	= 10 lb.moles/hr.
Reflux for distillation = 310.5 × 2.94	= 914 lb.moles/hr.
Total top tray reflux	= 924 lb.moles/hr.
Nitrogen entering reboiler = 924/0.94	= 983 lb.moles/hr.
Warm-end loss, exchanger A = (5.4 − 2.4) ° F. × 6.96 × 310.5	= 6,500 B.t.u./hr.
Warm-end loss, exchanger B (not including expander flow) = (983 − 310.5) × (5.3 − 2.4) × 6.96	= 13,600 B.t.u./hr.
Total losses (not including expander flow)	= 164,900 B.t.u./hr.
Expander flow = 164,900/(420 − 20)	= 413 lb.moles/hr.
Nitrogen compressor flow = 983 + 413	= 1,396 lb.moles/hr.

TABLE 5—OPTIMUM NUMBER OF THEORETICAL PLATES 90% RECOVERY
Basis: 1 lb.mole of 95% O₂ product/hour

Number of theoretical plates	13	15	18	22	27
Top tray reflux, lb.moles/hr.	3.26	2.99	2.91	2.86	2.84
Top tray vapor, lb.moles/hr.[1]	7.29	7.02	6.94	6.89	6.87
Installed cost of trays, $[2]	109.20	131.70	156.00	189.50	232.00
Reboiler nitrogen flow, lb.moles/hr.[3]	3.47	3.18	3.10	3.04	3.02
Compressor power, hp.	4.89	4.48	4.37	4.29	4.26
Installed cost of compressor, $	322.00	294.50	288.00	282.00	280.00
Installed cost of reboiler, $[4]	51.70	47.40	46.20	45.30	45.50
Fixed charges on compressor, $/yr.	80.50	73.62	72.00	70.50	70.00
Fixed charges on reboiler, $/yr.	12.93	11.85	11.55	11.33	10.13
Fixed charges on trays, $/yr.	27.30	32.93	39.00	47.38	58.00
Cost of power, $/yr.	169.00	154.60	151.00	148.20	147.10
Cost of exchanger B and warm-end loss, $/yr.[5]..	61.40	56.20	54.80	53.70	53.40
Total yearly cost, $	351.13	329.41	328.35	331.11	338.63

[1] Air flow + reflux flow − oxygen flow = top tray vapor.
[2] Average cost per theoretical plate is $1.25/(lb.mole)(hr. top tray vapor), including costs of insulation, cold box, structural steel and erection.
[3] Includes 6% allowance for flashing at reflux throttle valve.
[4] From Table 4, $14.90/(lb.mole)(hour of reboiler nitrogen flow).
[5] From Table 3, $17.69 yearly cost for 1.0 lb.mole/hr., based upon loading of exchanger B at 60 lb.moles/(hr.)(section).

TABLE 6.—OPTIMUM RECOVERY

Basis: 1 lb.mole of 95% O₂ product/hour

Recovery of oxygen, %	90	95	97	98.5	99.5
lb.moles of air/lb.mole of product	5.03	4.76	4.66	4.59	4.54
Yearly cost of supplying air, $ *	378.50	358.00	350.60	345.60	342.00
Yearly cost of column and reflux †	328.10	329.50	333.00	336.10	341.30
Total yearly cost	706.60	687.50	683.60	681.70	683.30

* $75.26 yearly cost/(lb.mole of air)(hr.). *See text.*
† Minimum yearly costs from Figure 7.

TABLE 7.—UTILITY REQUIREMENTS¹

120 ton/day

	hp.	Steam, lb./hr.	Electricity, kw.	Water, gal./min.
Nitrogen compressor	1,970	19,700	...	100
Air blower	435	4,350	...	60
Nitrogen exhaust blower	27	...	21	20
Nitrogen intercooler and aftercooler	...	...	...	530
Barometric condenser	...	...	...	1300
Scrubbing tower	...	...	...	360
Refrigeration unit	80	...	64	180
Water pumps	8	...	6	...
Regeneration blower²	14	...	11	...
Regeneration heater³	...	750	...	...
Regeneration cooler³	...	...	...	240
Instruments and valves	2	...	2	...
Expander⁴	—58	—580	1	5
	2,478	24,220	105	2795

¹ Based upon summer conditions.
² A steam rate of 10 lb./hp.hr. is assumed.
³ Load averaged over 24-hr. day.
⁴ 85% of expander power credited.

A summary of the utility requirements for this cycle is given in Table 7. As a matter of interest, a layout of the 120-ton plant is included as Figure 9.

Costs of Oxygen Production

On the basis of the utility requirements given in Table 7, the cost of oxygen production for the 120-ton plant is given in Table 8. The first calculation gives the oxygen production cost based on the fixed charge rate of 25%, the rate which has been used throughout this paper, and which is in the range often applied in the chemical and petroleum industries. Also included in Table 8 is a calculation of oxygen cost at 12.5% fixed charges, a rate which is frequently used in such industries as steel and manufactured gas. Since this last calculation makes use of operating and investment costs resulting from economic balances of the cycle variables for the 25% fixed charge rate, the oxygen cost of $6.80/ton is slightly higher than would be obtained in the most economical plant designed for the 12.5% fixed charge rate.

Similar estimates have been made to extend the range of plant capacities up to 1000 tons/day, and the results are plotted in Figure 10. While the oxygen costs on the curve for 12.5% fixed charges will again be slightly high, these figures are believed to be adequate for most estimating purposes. Any actual installation has, of course, to be considered in terms of its particular location, available utilities, and other special requirements.

General Remarks

Potential users of tonnage oxygen, when exploring the various processes, often ask the question, "How much power does the process require per ton of oxygen?" While the power requirement given in Table 7 may be typical of a 120-ton plant for many locations in the United States, it is still true that a large change in power cost or in the rate of fixed charges will have a marked effect on the choice of design conditions. An idea of the possible variations in power may be obtained from the fact that the change in compressor discharge pressure in Table 4 from 71 lb./sq.in. abs. to 115 lb./sq.in.abs. increases the compressor power by about 10%. Similarly, a change in warm-end temperature loss of 50% results in a change in compressor power by about 15%. Variations of such magnitude are en-

tirely possible, depending on the particular economic situation.

These differences are particularly noticeable when comparing information on European oxygen plants with American plants. It is apparently true that the relative magnitude of the value of power to the value of equipment is considerably greater in Europe than in the United States, so that in Europe, great emphasis is placed on economy of power. This results, in oxygen plant cycles, in lower compressor discharge pressures (4.5 to 5.5 atm. abs.), in lower warm-end temperature differences (3.5 to 7° F.), in lower pressure drops, and in higher compressor over-all efficiencies. Thus, the stated power consumption of European oxygen plants may be as much as 30% lower than figures quoted for American plants of comparable capacity. A paper by Shuftan (11) is a good example of the European approach, and it is interesting to compare both his cycle assumptions and his final power figures with the results given in the present paper. It should be noted, however, that in his discussion of the Rescol cycle, Shuftan uses a reflux rate which, according to our calculations given in Figure 6 of the present paper, is too low, after allowance is made for flashing. It may be that Shuftan has neglected to make an allowance for flashing at the reflux throttle valve.

In a recent comparison of power requirements of oxygen processes, Bliss and Dodge (2) use a wide variety of assumptions, which the authors state introduces an uncertainty of 10-20% in their power comparisons. They also seem to neglect such items as switching losses in regenerators, losses in acetylene removal devices, influence of argon in the column, and the effect of heat leak and other losses on oxygen recovery in double column plants. In the case of their cycle 7-II-A, which resembles the Elliott cycle, it is difficult to see the nature of the assumption which would cause 60% of the nitrogen to flow to

TABLE 8.—COST OF 95% OXYGEN *

120 tons/day

Item	Quantity	Unit Cost	Yearly Cost
Steam	24,220 lb./hr.	40 cents/M lb.	$83,800
Electricity	105 kw.	0.5 cents/kw.hr.	4,500
Water	2,795 gal./min.	1.0 cents/M gal.	14,500
Labor	2 men/shift	$1.60/hr.	27,600
Subtotal			$130,400
Fixed charges, 25% of investment †			325,000
Total			$455,400

Cost/ton of 95% oxygen = $455,400/4320 = $10.50

| Fixed charges, 12.5% of investment † | | | $162,500 |
| Total | | | $292,900 |

Cost/ton of 95% oxygen = $292,900/4320 = $6.80

* Based upon summer conditions.
† Includes depreciation, taxes, insurance and maintenance. Estimated turn-key cost of plant = $1,300,000.
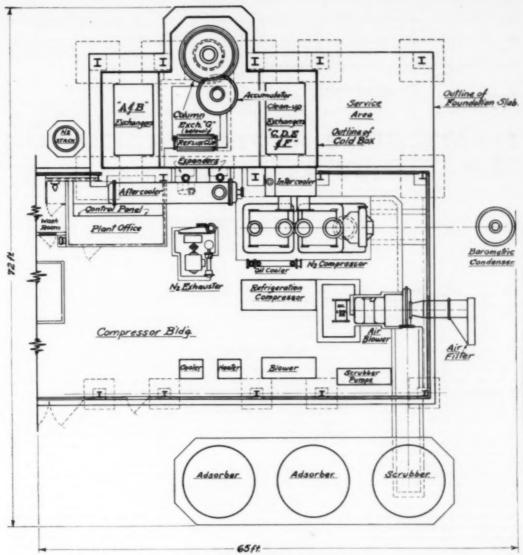
Fig. 9. Plan View of 120-Ton Plant.

the expander (their Table 8), while their other cycles require only 7.5 to 32% expander flow. This should be compared also with Figure 1 of the present paper which shows less than 30% expander flow. Whatever the nature of the assumption or error made by Bliss and Dodge, it seems likely that this is the cause of the unfavorable power comparison found for cycle 7-11-A.*

Another problem in which economic balances play an important part is that of the production of enriched air. For a number of applications of oxygen, particularly in the steel industry, the purity of oxygen desired is considerably lower than 95%. Such low purity oxygen may be made directly in an enriched air plant, or may be made by the mixing of higher purity oxygen with

* The writer wishes to take this opportunity to note that Bliss and Dodge are also in error in stating that the 4-6% of high pressure air used in the Linde-Fränkl cycle does not result in unbalanced flow in the regenerators. These authors reach this conclusion by calculating that at least an equivalent amount of nitrogen must be used for cooling of the high pressure side-stream. As may be checked with T-S diagrams, this is incorrect for an enthalpy balance based on the side-stream entering the column as a saturated vapor after throttling. In fact, there is no reason for cooling the air even this far other than economy of refrigeration, which, however, is of minor importance compared with the ability to clean up the regenerators successfully. This is confirmed by an early description of the Linde-Fränkl process (7), which states that the nitrogen used for cooling of the high pressure air is only ¾ of the quantity of air.

Fig. 10. Estimated Investment and Production Costs.

additional air. For example, the production of 1.0 lb.mole of 30% oxygen in an enriched air plant would require the cooling and distillation of about 1.45 lb.moles of air. On the other hand, 1.0 lb.mole of 30% oxygen may be obtained by the mixing of 0.878 lb.mole of normal air with 0.122 lb.mole of 95% oxygen. Production of this 95% oxygen would require the low temperature processing of about 0.56 lb.mole of air, which would represent a considerably lower investment cost than the enriched air plant, but which would theoretically require somewhat greater power consumption.

In a broader sense, the problem becomes that of determining, from a balance of fixed charges against the cost of power, the optimum oxygen purity to be produced in the low temperature plant, for admixture with additional air. In a recent economic study of this problem for Linde-Fränkl plants under European conditions, Karwat (8) concludes that the optimum purity is in the neighborhood of 73%. Converting this conclusion for the higher ratio of capital cost to power cost which prevails in the United States, it is apparent that the optimum purity will be appreciably higher. For the Elliott system, rough estimates indicate that the optimum purity will be not far from 95%, but this, again, must be confirmed for the particular economic conditions applying to each installation.

Acknowledgment

The writer wishes to express his appreciation to Ann Kindelan, who carried out most of the column calculations, and to Messrs. L. C. Claitor, J. L. Cost, D. B. Crawford, D. Aronson and M. Atcheson, all of whom have contributed at various times to the economic studies of the Elliott oxygen system.

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Discussion

C. W. Perry (Phillips Chemical Co., Bartlesville, Okla.): In previous studies, it was thought that perhaps this process, namely, the Elliott, would take up to 25% more power compared with the best of the Linde-Fränkl variations. This was partly because of possible smaller warm-end temperature differences in the Linde-Fränkl regenerative-type heat exchangers compared with the indirect Elliott ones. However, when all factors are taken into account, including the switching losses every two or three minutes in the Linde-Fränkl-type processes compared with one hour or more for the Elliott process, it was felt that the power requirements would compare favorably. The Elliott process is expected to give maximum operability and on-stream time. The method of producing column reflux gives more flexibility to the Elliott process than others.

The opportunities for sale of such plants in the synthesis of liquid fuels, oxygenated chemicals, iron and steel, city gas, etc., have been pointed out adequately in the literature. Synthetic liquid fuels from natural gas are not going into production as rapidly as was expected two or three years ago for basic reasons such as heavy commitments of gas reserves, heavy investment, unsolved engineering problems and the loss of British thermal unit value in converting natural gas to liquid fuel. However, future synthetic fuel developments based on coal will probably bring tonnage oxygen in that field up to expectancies. In the steel industry, certainly there are enough developments under way in three or four different large companies to indicate what the real value is. It has been indicated that increases in capacity in the open-hearth or blast-furnace parts of the plants require investments to meet that increase in other parts of the plant, which apparently has some sort of a braking effect. With respect to oxygenated chemicals, currently these seem to be in adequate supply, and that development may be slowed up. It is my opinion, however, that the erection of tonnage oxygen plants is in an incubation period at the present time and that ultimately, they will come up to the original expectancies.

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EVALUATION OF EQUILIBRIUM STAGES AND TRANSFER UNITS

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The first part of this paper presents a simple correlation for the calculation of the number of transfer units, (N.T.U.), at finite reflux, for systems with constant relative volatility in a packed fractionating column. A rapid graphical method replaces the tedious analytical solution with a maximum deviation of 6%. A close similarity is found when the new correlation is compared with two existing correlations developed for plate columns. The precise relationship and analogy between the plate column and the packed column are considered.

The latter part of this paper presents a simple type of equation, involving only two constants to represent the vapor-liquid equilibrium relationship for many nonideal systems. An equation is also developed for the determination of the minimum number of equilibrium stages which shows close agreement with the graphical solution. Successful application is also made of a new equation to determine the minimum number of transfer units.

CALCULATION of the number of transfer units in a packed column or the number of equilibrium stages in a plate column has been solved by various graphical (3, 12, 15, 22) and analytical methods (6, 7, 9, 11, 16, 17, 19). The former method provides a more general solution for all systems while the latter method applies only to systems where the relative volatility, α, is assumed to be a constant. Several improvements have been proposed (1, 4, 5, 13, 21) in the graphical determination of the number of transfer units. There are some empirical correlations (3, 10) for the calculation of the number of equilibrium stages in the plate columns. Nomographs (18, 20) and quadrographs (8) have been developed to simplify these calculations. A sound correlation, however, is still lacking for the calculation of the number of transfer units.

A simplified method is, therefore, proposed for the calculation of (N.T.U.) and equilibrium stages in packed and plate columns respectively, based on the proposed empirical relationships. In the present discussion, we shall divide the systems into constant or variable relative volatility.

Part 1: Systems with Constant Relative Volatility

A. Correlation for Determination of Number of Transfer Units (N.T.U.).

The number of transfer units in a packed fractionating column, if the gas phase resistance is controlling, is given as follows:

$$(N.T.U.) = \int_{y_1}^{y_2} \frac{dy}{y^* - y} \quad (1)$$

Under the condition of total reflux and constant relative volatility, Equation (1) has been integrated (3) to give:

$$(N.T.U.)_m = \frac{1}{\alpha - 1}\left[ln\left(\frac{x_D}{x_w}\right) + \alpha\, ln\left(\frac{1 - x_w}{1 - x_D}\right)\right] \quad (2)$$

For the condition of finite reflux, Equation (1) has been integrated (6, 5) for enriching and stripping sections respectively when a saturated liquid is used. The Dodge-Huffman equation (6), as applied under these conditions, is written as follows:

$$(N.T.U.) =$$

$$\frac{2R + B}{2\sqrt{B^2 + 4Ax_D}} ln \frac{(2Ax_D + B - \sqrt{B^2 + 4Ax_D})(2Ax_F + B + \sqrt{B^2 + 4Ax_D})}{(2Ax_F + B - \sqrt{B^2 + 4Ax_D})(2Ax_D + B + \sqrt{B^2 + 4Ax_D})}$$

$$+ \frac{2R' + B'}{2\sqrt{B'^2 - 4A'x_w}}$$

$$ln \frac{(2A'x_F + B' - \sqrt{B'^2 - 4A'x_w})(2A'x_F + B' + \sqrt{B'^2 - 4A'x_w})}{(2A'x_w + B' - \sqrt{B'^2 - 4A'x_w})(2A'x_w + B' + \sqrt{B'^2 - 4A'x_w})}$$

$$+ \frac{1}{2} ln \frac{Ax_F^2 + Bx_F - x_D}{Ax_D^2 + Bx_D - x_D} + \frac{1}{2} ln \frac{A'x_w^2 + B'x_w + x_w}{A'x_F^2 + B'x_F + x_w} \quad (3)$$
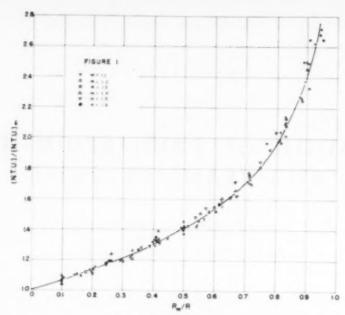
Fig. 1. N.T.U. Correlation.

where

$$A = R(1 - a)$$
$$B = R(a - 1) - x_D(a - 1) + a$$
$$R' = \frac{F - R}{F - 1}$$
$$A' = R'(1 - a)$$
$$B' = R'(a - 1) + x_w(a - 1) - a$$

Such integrated equations are complicated and their solution is tedious and time-consuming.

In order to develop a useful correlation in place of the tedious analytical calculation, it is found that the number of transfer units at finite reflux is a function of the reflux ratio R, the minimum reflux ratio R_m (involving variables a, x_F and x_D) and the number of transfer units under total reflux (involving variables a, x_D and x_w). Data

are correlated by plotting $(N.T.U.)/(N.T.U.)_m$ vs. R_m/R, as shown in Figure 1. This correlation is based on the following range of variables:

1. a from 1.1 to 1.6
2. x_F from 0.3 to 0.6
3. x_D from 0.70 to 0.99
4. x_w from 0.01 to 0.30
5. R from 3 to 200
6. R_m from 2.47 to 28.4
7. $(N.T.U.)_m$ from 8.43 to 53.7
8. $(N.T.U.)$ from 9.09 to 155

Approximately 150 points representing the systems of constant a are involved in this correlation. These calculations were done algebraically using Equations (2) and (3) for total and finite reflux respectively. The maximum deviation is about 6%, and the average deviation is

undoubtedly within the accuracy of the calculation. A comparison between the calculated and graphically determined values using Figure 1, is shown in Table 1.

Since this correlation is intended to simplify the calculation when the number of transfer units is large, a condition necessitating use of the analytical method, no attempt has been made to correlate the data where the number of transfer units is less than nine.

B. Comparison of Correlations Between Packed and Plate Columns. Correlations for plate columns similar to that shown in Figure 1 have been proposed by Brown and Martin (2) and also by Gilliland (10). For purposes of comparison the ratio J/J_m in the former calculation is replaced by R/R_m when the feed is saturated liquid. The correlation of Brown and Martin (for $n > 8$) is reproduced in Figure 2. Likewise, Gilliland's correlation has been replotted in a somewhat different manner by assuming:

$$\frac{n - n_m}{n + 1} \doteq 1 - \frac{n_m}{n}$$

when $n \gg 1$

$$\frac{R - R_m}{R + 1} \doteq 1 - \frac{R_m}{R}$$

when $R \gg 1$

TABLE 1—COMPARISON BETWEEN (N.T.U.) CALCULATED AND VALUES FROM FIGURE 1

a	x_F†	x_D	x_w	R_m	R	$(N.T.U.)_m$	$(N.T.U.)$ Calculated	$(N.T.U.)$ From Fig. 1	% Deviation
1.1	0.3	0.90	0.05	28.4	30.0	53.7	155.0	148.0	−4.5
1.1	0.4	0.80	0.01	16.7	18.0	23.6	58.8	58.0	−1.4
1.2	0.5	0.99	0.01	9.8	11.76	50.6	101.0	105.0	+4.0
1.3	0.5	0.95	0.05	9.0	14.4	32.4	50.5	51.0	+1.0
1.3	0.6	0.95	0.05	4.72	5.38	22.6	51.4	51.5	+0.2
1.3	0.5	0.80	0.20	3.6	6.0	19.6	16.5	16.5	0.0
1.4	0.5	0.95	0.05	4.4	6.16	17.7	31.1	30.6	−1.6
1.4	0.5	0.90	0.10	3.8	9.12	13.2	17.7	17.4	−1.2
1.5	0.5	0.99	0.01	3.9	5.0	23.0	44.1	43.7	−0.9
1.5	0.4	0.90	0.10	4.0	16.0	11.0	13.0	13.0	0.0
1.6	0.4	0.99	0.01	4.04	5.0	19.8	39.3	40.0	+1.8
1.6	0.5	0.90	0.10	2.47	10.0	9.5	11.3	11.3	0.0

* Saturated liquid feed.
† Calculated by Equation (3).

No notable error is found in these assumptions when n is greater than 10. From a study of Figure 2 a maximum deviation of about 10% is found between Brown and Martin's and the approximate Gilliland correlations over the range of R_m/R from 0 to 0.95. The correlation of $N.T.U.$ for the packed column, as presented in this paper, agrees closely with that of Brown and Martin for the plate column with a maximum deviation of 5% over the range of R_m/R from 0 to 0.85. This shows the close similarity between distillation in plate and in packed columns.

C. Relationship between Packed and Plate Columns. The mathematical relationship between the two types is readily derived under the condition of total reflux. The minimum number of equilibrium stages for a plate column developed by Fenske (9) is given as:

$$n_m = \frac{1}{ln\ a}\ ln\left(\frac{x_D}{x_w}\right)\left(\frac{1-x_w}{1-x_D}\right) \quad (4)$$

By the expansion of a logarithmic series, it is found that for values of a between 1.0 and 1.6, the following approximation can be made with an error of not greater than 2%:

$$ln\ a = 2\left(\frac{a-1}{a+1}\right)$$

Combining this approximation with Equations (2) and (4), one obtains

$$(N.T.U.)_m = \frac{2a}{a+1}\ (n)_m + ln\left(\frac{x_D}{x_w}\right) \quad (5)$$

or

$$\frac{(N.T.U.)_m}{(n)_m} = \frac{2a}{a+1} - \left(\frac{a-1}{a+1}\right)\left(\frac{1}{k}\right) \quad (6)$$

where

$$k = \frac{1}{2}\left[1 + \frac{ln\left(\frac{1-x_w}{1-x_D}\right)}{ln\left(\frac{x_D}{x_w}\right)}\right]$$

From the close similarity between the correlations for both packed and plate columns, as discussed in section B, one may conclude that for values of R_m/R not greater than 0.85, the following relationship is empirically true:

$$\frac{(N.T.U.)}{(N.T.U.)_m} = \frac{n}{n_m} \quad (7)$$

Substituting Equation (7) in Equation (6) the "$N.T.U.$ equivalent to a theoretical plate" is obtained from the following equation:

$$\frac{(N.T.U.)}{n} = \frac{2a}{a+1} - \left(\frac{a-1}{a+1}\right)\left(\frac{1}{k}\right) \quad (8)$$
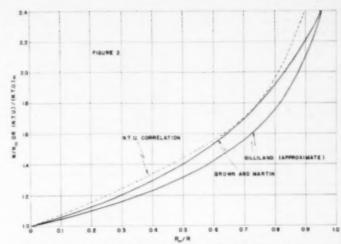
Fig. 2. Comparison of Gilliland, Brown, Martin and Present N.T.U. Correlations.

Equations (6) and (8) reveal two facts. First, the ratio of $(N.T.U.)$ and n is not only a function of a but also of the terminal conditions of separation, as defined by k. Second, when the separation is nearly complete or when a is close to unity, the number of equilibrium stages will be substantially equal to the number of transfer units, as shown by Dodge and Huffman's calculation (6). The following criteria can, therefore, be written:

Condition	Criteria
$N.T.U. = n$; $k = 1$ or $\left(\frac{1-x_w}{1-x_D}\right) = \left(\frac{x_D}{x_w}\right)$	
$N.T.U. > n$; $k > 1$ or $\left(\frac{1-x_w}{1-x_D}\right) > \left(\frac{x_D}{x_w}\right)$	
$N.T.U. < n$; $k < 1$ or $\left(\frac{1-x_w}{1-x_D}\right) < \left(\frac{x_D}{x_w}\right)$	

Part II: Non-Ideal Systems

A. Equilibrium Relationship. Formulation of a basic equation for the vapor-liquid equilibrium relationship for nonideal systems is necessary before any developments can be made. It is recognized that, while empirical constants are inevitably involved, their number should be reduced to a minimum. A power series, for example, can be used as shown by Eshaya (7), but the number of constants should be at least four in order to obtain a fairly good agreement. A formula has been derived empirically by Yu and Coull

(23) to represent the phase equilibrium relationship with not more than four constants. For highly nonideal systems encountered in distillation, such as the ethanol-water system, the following relationship can be used:

$$\left(\frac{y}{1-y}\right) = a\left(\frac{x}{1-x}\right) Exp.\left\{b + c\ ln\left(\frac{x}{1-x}\right)\left[ln\left(\frac{x}{1-x}\right) + d\right]\right\} \quad (9)$$

where y and x represent the vapor and liquid composition respectively. Equation (9), although originally used at constant temperature, can be used under the condition of constant pressure. For moderately nonideal systems, like those shown in Table 2, the following relationship with only two constants is sufficient to represent the data with a maximum deviation of less than 2% over a concentration range of not less than 0.04 to 0.96 mole fraction.

$$\left(\frac{y}{1-y}\right) = a\left(\frac{x}{1-x}\right)^b \quad (10)$$

A detailed discussion of Equations (9) and (10) is beyond the scope of this paper.

The advantage of using Equations (9) and (10) in preference to the power series can be seen by the following examples. Eshaya (7) uses the

following expression for the system chloroform-benzene with a maximum deviation of 4.5%.

$$y = 1.30x + 0.25x^2 - 0.54x^3$$

In this case, Equation (10) is sufficient to represent the data with a maximum deviation of 1.5%.

$$\left(\frac{y}{1-y}\right) = 1.92\left(\frac{x}{1-x}\right)^{1.17}$$

For azeotropic systems Equation (10) can be used with great convenience. A solution of Equation (10) for the condition where $y = x$ shows the following two roots with the third root extrapolated:

$$\begin{cases} y_1 = 0 \\ y_2 = \dfrac{1}{a^{\left(\frac{1}{b-1}\right)} + 1} \\ y_3 = 1 \text{ (by extrapolation)} \end{cases}$$

The second root gives the azeotropic composition. For the system in which no azeotrope is formed, the second root reduces closely to zero (if $b > 1$) or unity (if $b < 1$) as shown in Figure 3. The existence of y_2, therefore, is the limiting condition for Equation (10) when it is applied to a system without an azeotrope.

Systems evaluated according to Equation (10) are shown in Table 2. In this table the values of a and b are so determined that the maximum accuracy is found in the computation of $(N.T.U.)$ or n for nearly complete separation because only under this condition is the algebraical solution desirable. This means that the values 3.55 and 0.81 for a and b respectively for the methanol-water system may not be the best choice over a range of from 0.1 to 0.9 but will represent the best choice from 0.9 to 1.0 mole fraction because of the nature of the equilibrium curve.

Equation (10) reduces to the equation of constant volatility if constant b is equal to unity.
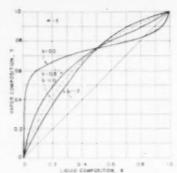
Fig. 3. Graphical Representation of Eq. (10)

B. Calculation of Minimum Number of Equilibrium Stages for a Plate Column. For all separation processes the minimum number of stages is found at total reflux. The derivation of a general expression is possible where Equation (10) is used to represent the equilibrium relationship.

From Equation (10) one obtains for the first plate from the top,

$$\left(\frac{x_D}{1-x_D}\right) = a\left(\frac{x_1}{1-x_1}\right)^b$$

Since at total reflux, the composition of vapor ascending from the second plate should be the same as that of the liquid overflowing from the first plate, or

$$x_1 = y_1$$

Then for the second plate,

$$\left(\frac{x_1}{1-x_1}\right) = \left(\frac{y_1}{1-y_1}\right) = a\left(\frac{x_2}{1-x_2}\right)^b$$

which may be written as:

$$\left(\frac{x_D}{1-x_D}\right) = a\left[a\left(\frac{x_2}{1-x_2}\right)^b\right]^b$$

or in terms of n,

$$= a^{1+b}\left(\frac{x_2}{1-x_2}\right)^{b^2}$$

For the nth plate, one obtains:

$$\left(\frac{x_D}{1-x_D}\right) = a^{1+b+b^2+\ldots+b^{n-1}}\left(\frac{x_n}{1-x_n}\right)^{b^n} \tag{11}$$

There are three cases expressed by Equation (11) depending upon the value of b.

Case I: When b is equal to unity.

Equation (11) reduces to:

$$\left(\frac{x_D}{1-x_D}\right) = a^n\left(\frac{x_n}{1-x_n}\right)$$

which is identical with Fenske's equation for total reflux as applied to the system where the relative volatility is a constant.

Case II: Where b is greater or less than unity.

In this case the geometric series in the exponent can be evaluated when b is less than or greater than unity as follows:

$$1 + b + b^2 + \ldots + b^{n-1} = \frac{(1-b^n)}{1-b}$$

Equation (11), therefore, can be expressed as:

$$\left(\frac{x_D}{1-x_D}\right) = a^{\left(\frac{1-b^n}{1-b}\right)}\left(\frac{x_w}{1-x_w}\right)^{b^n} \tag{12}$$

Solving for n and replacing subscript n by w to represent the bottom composition, the following is obtained:

$$\left(\frac{x_D}{1-x_D}\right) = a^{\left(\frac{1}{1-b}\right)}\left[a^{-\left(\frac{1}{1-b}\right)}\left(\frac{x_w}{1-x_w}\right)\right]^{b^n}$$

$$n = \frac{1}{\ln b}\ln\left[\frac{\left(\frac{1}{b-1}\right)\ln a + \ln\left(\frac{x_D}{1-x_D}\right)}{\left(\frac{1}{b-1}\right)\ln a - \ln\left(\frac{1-x_w}{x_w}\right)}\right] \tag{13}$$

Equation (13) gives the relationship of the minimum number of equilibrium stages at the condition of total reflux for a nonideal system. It should be noted that Equation (13) is true only for the systems whose equilibrium relationship can be represented by Equation (10). Table 3 shows the result of applying Equation (13) compared with the actual graphical computation. The maximum deviation is found to be

TABLE 2.—CONSTANTS FOR LIQUID-VAPOR EQUILIBRIUM SYSTEMS

Pressure 1 atm

System (?)	a	b
Chloroform-benzene	1.92	1.17
Benzene-ethylene dichloride	1.11	1.00
Acetic acid-water	0.606	0.93
Methanol-water	3.55	0.81
Acetic acid-benzene	1.635	0.53
Acetone-ethanol	2.12	0.845
Acetone-methanol	1.35	0.78
Carbon tetrachloride-ethyl acetate	1.051	0.87
Carbon tetrachloride-toluene	2.46	1.00
Chloroform-acetone	0.844	1.273

2.7% which is within the accuracy of a graphical solution.

For a small number of highly non-ideal systems it is necessary to use Equation (9) for the equilibrium relationship. Although no general solution has been developed to date for highly nonideal systems, it has been found that the equilibrium curve for such systems can be divided into two portions and that each portion obeys Equation (10).

This technique has been applied to the system ethanol-water. The point of division is taken at 0.5 mole fraction. This treatment shows great advantage in computing the number of equilibrium stages near the azeotropic composition. In the interest of conserving space this treatment is omitted but the numerical table will be sent to the reader upon request.

C. Calculation of Minimum Number of Transfer Units. Theoretically, the minimum number of transfer units can be found by the integration of Equation (1) upon substitution of Equation (10). By means of the binomial series, the expressions x^b and $(1-x)^b$ are expanded and the integration of Equation (1) is mathematically possible. The final result is tedious and has no value in application. For all practical purposes a method similar to that of finite difference may be used to simplify the calculation.

Assuming over the concentration range of one equilibrium stage (one theoretical plate) the relative volatility to be approximately constant. This assumption is reasonably true especially when the average relative volatility closely approaches unity or the concentration range over an equilibrium stage is small. Let x_w and x_1 represent the concentration range over the first plate from the bottom, then the number of transfer units covering this range can be calculated by Equation (5) by setting $n=1$; therefore,

$$(N.T.U.)_1 = -ln\left(\frac{x_1}{x_w}\right) + \frac{2a_w}{a_w+1}$$

where a_w represents the average relative volatility over the first plate from the bottom. By Equation (10) since a_w is assumed to be a constant over one plate, one can set:

$$a_w = a\left(\frac{x_w}{1-x_w}\right)^{b-1}$$

For the second plate from the bottom covering the concentration range of x_1 to x_2, one obtains

$$(N.T.U.)_2 = -ln\left(\frac{x_2}{x_1}\right) + \frac{2a_1}{a_1+1}$$

where

$$a_1 = a\left(\frac{x_1}{1-x_1}\right)^{b-1}$$
$$= a\left[a\left(\frac{x_w}{1-x_w}\right)^b\right]^{b-1}$$
$$= a_w^b$$

Similarly, for the $(n-1)th$ plate from the bottom, the following is obtained

$$(N.T.U.)_n = -ln\left(\frac{x_n}{x_{n-1}}\right) + \frac{2a_{n-1}}{a_{n-1}+1}$$

where

$$a_{n-1} = a_w^{b^{n-1}}$$

Let x_n represent the top composition x_D, then the total number of transfer units equivalent to n equilibrium stages will be:

$$(N.T.U.) = -ln\left(\frac{x_D}{x_w}\right) + 2\left[\frac{a_w}{a_w+1} + \frac{a_w^b}{a_w^b+1} + \frac{a_w^{b^2}}{a_w^{b^2}+1} + \cdots + \frac{a_w^{b^{n-1}}}{a_w^{b^{n-1}}+1}\right]$$

$$= -ln\left(\frac{x_D}{x_w}\right) + 2S \qquad (14)$$

where S represents the summation of the series. The exact mathematical solution has not yet been found. When the number of transfer units is small, Equation (14) can be solved by stepwise summation. When the number of transfer units is large or b is close to unity, the following approximation can be made:

$$S = n\left[\frac{a_w^{(1+b+b^2+\cdots+b^{n-1})\frac{1}{n}}}{a_w^{(1+b+b^2+\cdots+b^{n-1})\frac{1}{n}}+1}\right]$$

TABLE 3—COMPARISON BETWEEN GRAPHICAL AND ANALYTICAL SOLUTIONS

System: Chloroform-benzene (1 atm.)

$a = 1.92;\ b = 1.17$

x_D	x_w	n_m from Graphical Method	n_m from Eq. (13)	Δn_m	a_w from Eq. (15)	$(N.T.U.)_m$ from Graphical Method	$(N.T.U.)_m$ from Eq. (15)	$\Delta(N.T.U.)_m$
0.95	0.05	13.00	12.87	—0.13	1.560	12.78	13.10	+0.32
0.90	0.05	12.40	12.12	—0.18	1.528	11.40	12.00	+0.60
0.85	0.05	11.60	11.60	0.0	1.496	10.76	10.27	—0.49
0.595	0.05	10.00	9.86	—0.14	1.402	8.22	9.10	+0.88
0.95	0.10	9.20	9.04	—0.16	1.767	9.50	9.60	+0.10
0.90	0.10	8.50	8.30	—0.20	1.699	8.13	8.50	+0.37
0.85	0.10	8.00	7.80	—0.20	1.646	7.49	7.88	+0.39
0.95	0.133	8.00	7.84	—0.16	1.926	8.51	8.68	+0.17
0.95	0.234	6.00	6.10	+0.10	1.970	6.92	6.98	+0.06
0.80	0.20	4.90	4.82	—0.08	1.777	4.33	5.01	+0.68

System: Methanol-water

$a = 3.55;\ b = 0.81$

x_D	x_w	n_m from Graphical Method	n_m from Eq. (13)	Δn_m	a_w from Eq. (15)	$(N.T.U.)_m$ from Graphical Method	$(N.T.U.)_m$ from Eq. (15)	$\Delta(N.T.U.)_m$
0.99	0.01	8.00	8.05	+0.05	3.13	8.47	8.92	+0.45
0.99	0.05	7.38	7.30	—0.08	2.81	8.05	8.71	+0.66
0.99	0.10	6.90	6.90	0.0	2.67	7.87	8.55	+0.68
0.95	0.01	5.40	5.26	—0.14	4.20	5.48	5.34	—0.14
0.90	0.20	4.00	4.06	+0.06	4.20	4.45	4.20	—0.25
0.80	0.01	3.70	3.60	—0.10	5.28	3.20	3.00	—0.20
0.90	0.10	3.27	3.26	—0.01	3.85	3.90	3.81	—0.09
0.85	0.10	2.81	2.78	—0.03	4.12	3.20	3.03	—0.17
0.80	0.20	1.95	2.00	+0.05	4.00	2.52	2.28	—0.24

It is seen that when b is equal to unity, Equation (14) reduces to Equation (5) for the case where a is a constant. By the summation of the geometric series in the exponent, one obtains

$$S = n \left[\frac{a_w \left(\frac{1-b^n}{1-b} \right) \frac{1}{n}}{a_w \left(\frac{1-b^n}{1-b} \right) \frac{1}{n} + 1} \right]$$

An average volatility, a_{av}, is defined so as to give the same number of equilibrium stages from x_D to x_w, according to Equation (12), then,

$$\left(\frac{x_D}{1-x_D} \right) = a_{av}{}^n \left(\frac{x_w}{1-x_w} \right)$$

On comparison with Equation (12), it is obvious that:

$$a_{av} = \left[a \left(\frac{1-b^n}{1-b} \right) \left(\frac{x_w}{1-x_w} \right)^{b^n-1} \right]^{\frac{1}{n}}$$

$$= \left[a \left(\frac{x_w}{1-x_w} \right)^{b-1} \right] \left(\frac{1-b^n}{1-b} \right) \frac{1}{n}$$

$$= a_w \left(\frac{1-b^n}{1-b} \right) \frac{1}{n}$$

The following equation is, therefore, obtained:

$$(N.T.U.) = - ln \left(\frac{x_D}{x_w} \right) + \frac{2na_{av}}{a_{av} + 1}$$

(15)

where

$$a_{av} = \left[\left(\frac{x_D}{1-x_D} \right) \left(\frac{1-x_w}{x_w} \right) \right]^{\frac{1}{n}}$$

Equation (15) means that the average relative volatility which gives the actual number of equilibrium stages will be the average relative volatility which gives the actual number of transfer units under the same terminal condition. The term n is evaluated by means of Equation (13) before the evaluation of $(N.T.U.)$. The result is shown and compared in Table 3. From this table, it is seen that the value of average relative volatility changes sharply with the composition. It is remarkable to note that in the case of different terminal conditions but with the same number of equilibrium plates at total reflux, both a_{av} and the number of transfer units could be widely different. From Equation (15) it is possible to obtain a wide difference, as shown in Table 3. In general, the evaluation of the number of transfer units by Equation (15) is not as good as that for the number of equilibrium plates by Equation (13). This is due to the inaccurate approximation

of Equation (14) and the summation of the series, S.

Conclusions

A simple correlation for the calculation of the number of transfer units in a packed column for systems whose relative volatility is a constant is presented based on approximately 150 points. Comparison of the previous correlations which deal only with plate columns has been made. The fundamental relationship between the number of transfer units and the number of equilibrium plates is derived and discussed.

For systems whose relative volatility is not a constant, a simple equilibrium relationship is presented involving only two constants. This relationship is shown to be useful in many systems with a maximum deviation of less than 2%. Equations for the determination of the number of equilibrium plates and transfer units at total reflux have been presented based on the proposed equilibrium relationship. The calculated result shows good agreement with the ordinary graphical method.

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Notation

a,b,c,d = empirical constants

F = feed rate

J = symbol used by Brown and Martin (16)

$(N.T.U.)$ = number of transfer units

n = number of equilibrium stages

R = reflux ratio, lb. moles of reflux/lb. mole of top product

S = summation of the series involved in Equation (14)

x = liquid composition, mole fraction

y = vapor composition, mole fraction

y^* = vapor composition, mole fraction, in equilibrium with liquid composition x

a = relative volatility

SUBSCRIPTS:

$_{av}$ = average

$_D$ = top product

$_F$ = feed

$_m$ = minimum number of

$_n$ = nth individual condition

$_w$ = bottom product

$_{1,2,\ldots}$ = individual condition

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DIFFUSIONAL PROPERTIES OF MULTICOMPONENT GASES

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From the Stefan-Maxwell diffusion equations the following expression is derived for the effective diffusion coefficient of a gas with respect to a multicomponent mixture of stagnant gases:

$$D'_A = \frac{1 - y_A}{\dfrac{y_B}{D_{AB}} + \dfrac{y_C}{D_{AC}} + \dfrac{y_D}{D_{AD}} + \ldots}$$

Application of this relation is studied by comparison with exact solutions of the diffusion equations for various types of diffusion and is supported by experimental evidence. Application to the calculation of viscosities of gas mixtures is shown.

Approximate methods for calculation of rates of diffusion in complex mixtures for stagnant films, and also with simultaneous diffusion of all components are presented.

Effects of simultaneous diffusion on mass-transfer coefficients are discussed.

THE general interpretation of mass-transfer processes such as absorption, distillation and other diffusional operations is commonly based upon the well-known film theory of mass transfer. In this theory the complex process of interphase mass transfer in flow systems is reduced to an equivalent process of pure diffusion over an "equivalent film thickness" or effective length of diffusion path in the same fluid medium offering a resistance to diffusion equal to the total resistance of the actual system. In general the equivalent film thickness may be correlated as a function of the curvature of the liquid or solid surface over which the gas is flowing, of the total rate of flow and of the physical properties of the fluid stream in which the mass transfer is taking place. If general relations are available to predict the equivalent film thickness in a given apparatus it is possible to predict the rate of mass transfer in the system by application of the appropriate rate equations for pure diffusion.

Most experimental studies have been restricted to simple gas mixtures to which the equations for diffusion in binary gas mixtures are applicable. However, in the more general applica-

tion of mass-transfer correlations based upon such experiments it is necessary to consider systems of more than two components. It is therefore of considerable practical interest to consider the theory of diffusion in multicomponent mixtures.

Sherwood (24) has reviewed the general theory of diffusion in gases as developed by Maxwell (20) and Stefan (29, 30) and presents general differential equations applicable to any number of components. The integrations of these general equations for binary systems are well known. Gilliland (25) has solved the equations for simultaneous steady-state diffusion of two gases through a third stagnant gas. Integrations of the general equations for more than three components have not been available. Furthermore their solution would result in relations too complex and unwieldy for convenient use in engineering calculations.

It is the purpose of the present paper to develop an approximate method for calculation of steady-state diffusion in complex gas mixtures which is sufficiently accurate for most purposes and yet which avoids the mathematical difficulty of a rigorous treatment.

Effective Diffusion Coefficients in a Multicomponent Mixture

For the steady-state unidirectional diffusion of gas A through a second stagnant gas the following rate equation is applicable:

$$N_A = -\frac{D_{AB} P}{R T p_i}\left(\frac{dp_A}{dx}\right) \quad (1)$$

where:

N_A = rate of diffusion of component, A, (g. moles)/(sq. cm.) (sec.)

D_{AB} = diffusion coefficient for A in B sq.cm./sec.

P = total pressure, atm.

R = gas constant, (cu. cm.) (atm.)/(g.mole) (° K.)

T = absolute temperature, ° K.

p_A = partial pressure of component A, atm.

p_i = partial pressure of non-diffusing gas, atm.

x = distance in direction of diffusion, cm.

For diffusion of A into a complex mixture the analogous rate equation may be written:

$$N_A = -\frac{D'_A P}{R T p_i}\left(\frac{dp_A}{dx}\right) \quad (2)$$

where

D'_A represents the effective average diffusion coefficient of component A with respect to the composition of the gas.

For diffusion of gas A into a mixture of gases, $B, C, D \ldots$, according to the concepts of Maxwell:

$$-\frac{dp_A}{dx} = a_{AB}\,\frac{p'_A}{M_A}\,\frac{p'_B}{M_B}\,(U_A - U_B)$$

$$+ a_{AC}\,\frac{p'_A}{M_A}\,\frac{p'_C}{M_C}\,(U_A - U_C)$$

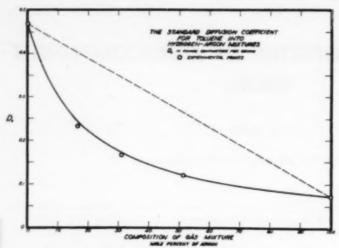$$+ a_{AD}\,\frac{p'_A}{M_A}\,\frac{p'_D}{M_D}\,(U_A - U_D) \ldots$$

$$(3)$$

Fig. 1. Comparison of Calculated Curve with Experimental Points for Diffusion of Toluene into Hydrogen-Argon Mixtures.

toluene into hydrogen-argon mixtures and ethyl propionate into hydrogen-air mixtures using the semi-infinite column technique suggested by Arnold (1). These experiments are to be described in detail in a subsequent paper. Results for the hydrogen-argon system corrected to 0° C., 1 atm. pressure are shown in Figure 1. The solid curve through the data is calculated by Equation (6) using the measured binary diffusivities. Marked deviation from a molal average relationship is illustrated by the shape of Figure 1 which is typical of curves calculated for three component systems. These experimental results are an encouraging indication of the validity of the Stefan-Maxwell equations which have been used to test the approximate calculation procedures developed in the present paper.

Equimolal Diffusion Plus Bulk Flow Concept. For convenience in mathematical analysis the transfer process for a given component may be viewed as a transfer of the material relative to a fixed plane of reference by the combined mechanisms of equimolal counter-diffusion and bulk flow of the total gas mixture. In the case of mass transfer through a gas film, bulk flow may result from a change in the total number of moles in the process, so that bulk flow does not necessarily mean that component gases pass through the film other than by diffusion. For example, in steady-state absorption there is a net transfer of gas through the film, or bulk flow, which is identical with and equal to the rate of diffusion of the absorbed component. This general viewpoint, although hypothetical in some instances, is useful in the development of rate equations for diffusion under conditions involving simultaneous transfer of all components of a mixture.

General differential equations for diffusion accompanied by bulk gas movement in binary systems have been derived by Stefan and Maxwell and alternatively by Jellinek (17) and Colburn and Hougen (8). To demonstrate this viewpoint Equation (3) may be written for a binary mixture and rearranged in a manner similar to the development of Equation (4), but noting that $P = p_A + p_B$ and U_B is not zero, to give the relation:

$$-\frac{dp_A}{dx} = \frac{RT}{D_{AB}P}(PN_A - p_A N_A - p_A N_B)$$

(7)

where N_B is the total rate of transfer of component B, (moles)/(sec.)(sq. cm.).

Introducing $y_A = \dfrac{p_A}{P}$ in Equation

where:

$a_{AB}, a_{AC} \ldots$ = proportionality constant for diffusion

$\rho'_A, \rho'_B \ldots$ = partial densities of components A, B, . . . g./cu.cm.

$M_A, M_B \ldots$ = molecular weight of A, B . . . g./mole

$U_A, U_B \ldots$ = average velocity of A, B . . . in the x direction, cm./sec.

In the case of all gases being stagnant except A; $U_B, U_C \ldots$ may be set equal to zero. Then introducing the relations:

$$a_{AB} = \frac{R^2 T^2}{D_{AB}P}, \quad a_{AC} = \frac{R^2 T^2}{D_{AC}P} \ldots$$

$$\rho'_A = \frac{p_A M_A}{RT}, \quad \rho'_B = \frac{p_B M_B}{RT} \ldots$$

$$U_A = \frac{M_A N_A}{\rho'_A}$$

Equation (3) reduces to the following expression:

$$-\frac{dp_A}{dx} = \frac{N_A RT}{P}\left[\frac{p_B}{D_{AB}} + \frac{p_C}{D_{AC}} + \frac{p_D}{D_{AD}} + \ldots\right]$$

(4)

Equation (4) may be rearranged in the following form:

$$N_A = -\frac{(1-y_A)P}{\left[\dfrac{y_B}{D_{AB}} + \dfrac{y_C}{D_{AC}} + \dfrac{y_D}{D_{AD}} + \ldots\right] RT p_t} \frac{dp_A}{dx} \ldots$$

(5)

where:

$y_A, y_B, etc.$ = mole fraction of components A, B, etc., in gas mixture

Comparison of Equations (2) and (5) indicates the following relation for the effective diffusion coefficient * for component A:

$$D'_A = \frac{1 - y_A}{\dfrac{y_B}{D_{AB}} + \dfrac{y_C}{D_{AC}} + \dfrac{y_D}{D_{AD}} + \ldots}$$

(6)

This equation may be written in the alternate form

$$D'_A = \frac{1}{\dfrac{y_{B'}}{D_{AB}} + \dfrac{y_{C'}}{D_{AC}} + \dfrac{y_{D'}}{D_{AD}} + \ldots}$$

(6a)

Where $y_{B'}, y_{C'} \ldots$ are the mole fractions of $B, C, \ldots$ in the inert gas portion considered separately from diffusing gas A. Thus for diffusion of A through a stagnant gas mixture, D'_A is constant for a given inert gas composition and independent of the mole fraction of A.

Fairbanks and the author (10) have verified Equation (6) for diffusion of

* This result was partially anticipated by Stefan (31).

(7) and rearranging gives the desired form:

$$N_A = -\frac{D_{AB}P}{RT}\frac{dy_A}{dx} + (N_A + N_B)y_A$$

(8)

The first half of the right side of (8) may be viewed as representing the rate of transport of A by equimolal diffusion, and the second half may be viewed as the rate at which A is carried along by bulk movement of the total gas phase.

For steady-state transfer of component A of a multicomponent system, the defining rate equation may be written by analogy to Equation (8)

$$N_A = -\frac{D'_A P}{RT}\left(\frac{dy_A}{dx}\right) + N_t y_A$$

(9)

where

N_t = total net rate of gas transfer[*] by diffusion and bulk flow, g.mole/(sec.)(sq.cm.)

D'_A = effective diffusion coefficient, sq.cm./sec.

For diffusion of A into a stagnant gas mixture, corresponding to $N_A = N_t$ in Equation (9), D'_A is given by Equation (6). The more general application and integration of Equation (9) for various types of diffusion will now be considered.

Film Pressure Factor Concept. The general type of diffusion calculation when two or more components are diffusing simultaneously may be conveniently handled by use of the film pressure factor concept in conjunction with a properly defined effective diffusion coefficient as developed below. The term, "film pressure factor," has been previously introduced by Hougen and Watson (16). The present development leads to a similar result for the film pressure factor term, although the derivation is different.

It is convenient to define a function ϕ for each component

$$\phi_A = \frac{N_A}{N_t}$$

(10)

$$\phi_B = \frac{N_B}{N_t}, \text{ etc.}$$

(11)

so that Equation (9) becomes

$$N_A = -\frac{D'_A P}{RT}\frac{dy_A}{dx} + \frac{N_A}{\phi_A}y_A$$

(12)

* N_t is the sum of rates of transfer for each component, with a rate of transfer in the $+x$ direction taken as positive and a rate of transfer in $-x$ direction taken as negative. For a binary system $N_t = N_A + N_B$.

Similarly for the diffusion of B

$$N_B = -\frac{D'_B P}{RT}\frac{dy_B}{dx} + \frac{N_B}{\phi_B}y_B$$

(13)

For steady-state diffusion, assuming a constant average diffusion coefficient, Equation (12) may be integrated to give[*]

$$N_A = \frac{\phi_A D''_A P}{RTx} \ln \frac{\phi_A - y_{A2}}{\phi_A - y_{A1}}$$

(14)

Introducing $y_A = \frac{p_A}{P}$ and rearranging

$$N_A = \frac{D''_A P}{RTx(p_f)_A}(p_{A1} - p_{A2})$$

(15)

where $(p_f)_A$ is the film pressure factor for diffusion of component A and is defined as:

$$(p_f)_A = \frac{P}{\phi_A}(\phi_A - y_A)_{lm} \quad (16)$$

and $(\phi_A - y_A)_{lm}$ is the log mean average over the boundary conditions of the film, i.e.,

$$(\phi_A - y_A)_{lm} = \frac{(\phi_A - y_{A2}) - (\phi_A - y_{A1})}{\ln \frac{\phi_A - y_{A2}}{\phi_A - y_{A1}}}$$

(17)

Similar equations may be written for diffusion of other components of the mixture. For component B

$$N_B = \frac{D''_B P}{RTx(p_f)_B}(p_{B1} - p_{B2})$$

(18)

and

$$(p_f)_B = \frac{P}{\phi_B}(\phi_B - y_B)_{lm} \quad (19)$$

For the special case of diffusion of one gas through a second stagnant gas

$$\phi_A = 1$$

and

$$(p_f)_A = (P - p_A)_{lm} = (p_i)_{lm}$$

so that Equation (15) reduces to the well-known equation

$$N_A = \frac{D''_A P(p_{A1} - p_{A2})}{RTx(p_i)_{lm}} \quad (20)$$

For equimolal counter diffusion $\phi_A = \infty$ and

$$(p_f)_A = P$$

This result is indicated directly by integration of Equation (9).

For a chemical reaction at a catalyst surface with diffusion of components to and from the interface the film pressure

* Deed, Schutz and Drew (4a) have proposed an equivalent expression for diffusion in binary systems.

factor may be expressed as follows: For the reaction:

$$aA + bB + \ldots = rR + sS + \ldots$$

where a, b, r, s are coefficients of the molecular species A, B, R, S undergoing reaction

$$\phi_A = \frac{a}{a + b - r - s}$$

and $(p_f)_A$ is given by Equation (16) as before. Similarly for component R

$$\phi_R = \frac{-r}{a + b - r - s}$$

The film pressure factor in the rate equation accounts for effects of bulk movement of the total gas mixture, due to simultaneous movement of the various components, on rate of transport of a given component under consideration. As shown below, it is necessary in some cases to take further into account the relative motions of the various components insofar as they may affect the value of the effective diffusion coefficient, D''_A, in Equation (15).

For diffusion of component A through a stagnant gas the film pressure factor is equal to the log mean partial pressure of the gases other than A over the diffusion path. If the other components are diffusing against A the film pressure factor will increase above the value for the stagnant gas case, thus decreasing the rate of diffusion of A. In cases of parallel diffusion the film pressure factor will be less than the value for the stagnant gas case. Thus the film pressure factor accounts for the opposing effect of components diffusing against the component under consideration and the carrying effect of components diffusing in the same direction. The principal justification for use of the film pressure term lies in its mathematical convenience and the resulting arrangement of the rate equation in a form similar to the more commonly used equations for equimolal counter-diffusion and the stagnant gas case.

Equations such as (15) and (16) may be written for any component of a mixture, and are the general equations which will be used to define the rates of diffusion. It now remains to determine the proper average effective diffusion coefficient D''_A for use in these equations.

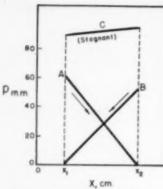*Solution of Stefan-Maxwell Equations for Three Components.* Before

Fig. 2. Setup for Test Calculation.

proceeding to the detailed treatment of approximate calculation procedures, the reference method used to test these approximate procedures will be described. Consider the simultaneous diffusion of gases A and B through a third stagnant gas C over an effective film thickness $x = x_2 - x_1$. Such a situation is approximated in fractional distillation columns where the process is one of nearly equimolal counterdiffusion. The partial pressures of the components at points 1 and 2 are maintained constant at any desired values under conditions of steady-state flow. Figure 2 shows partial pressure gradients for a typical test example at a total pressure of 150 mm. By proper adjustment of partial pressures rates of diffusion of A and B may be set over a range of desired values.

Sherwood (25) presents differential equations applicable to this case as follows:

$$\frac{-P}{RT}\frac{dp_A}{dx} = \left[\frac{1}{D_{AB}} - \frac{1}{D_{AC}}\right]N_A \cdot p_B - \left[\frac{N_B}{D_{AB}} + \frac{N_A}{D_{AC}}\right]p_A + \frac{N_A P}{D_{AC}} \quad (21)$$

$$\frac{-P}{RT}\frac{dp_B}{dx} = \left[\frac{1}{D_{AB}} - \frac{1}{D_{AC}}\right]N_B \cdot p_A - \left[\frac{N_A}{D_{AB}} + \frac{N_B}{D_{BC}}\right]p_B + \frac{N_B P}{D_{BC}} \quad (22)$$

Gilliland (25) has integrated (21) and (22) to give the two simultaneous equations:

$$\frac{N_A}{D_{AC}} + \frac{N_B}{D_{BC}} = \frac{P}{RTx}\ln\frac{p_{C_2}}{p_{C_1}} \quad (23)$$

$$N_A + N_B =$$

$$\frac{D_{AB}P}{RTx}\ln\left[\frac{\left[\frac{1}{D_{AB}} - \frac{1}{D_{AC}}\right]\frac{N_A+N_B}{N_B}p_{B_2} - \frac{N_A+N_B}{N_A}p_{A_2} + \left[\frac{\frac{1}{D_{AC}} - \frac{1}{D_{BC}}}{\frac{1}{D_{AB}} - \frac{1}{D_{BC}}}\right]P}{\left[\frac{1}{D_{AB}} - \frac{1}{D_{AC}}\right]\frac{N_A+N_B}{N_B}p_{B_1} - \frac{N_A+N_B}{N_A}p_{A_1} + \left[\frac{\frac{1}{D_{AC}} - \frac{1}{D_{BC}}}{\frac{1}{D_{AB}} - \frac{1}{D_{BC}}}\right]P}\right]$$

$$(24)$$

To serve as a standard for other calculation procedures a number of test examples were set up covering a wide range of types of diffusion from nearly equimolal to high relative rates of counterdiffusion and parallel diffusion among two components in the presence of a third stagnant component. The system hydrogen-water-vapor-carbon dioxide was selected for study to give the examples physical significance and to include a wide range of diffusional properties. The total pressure was arbitrarily set at 150 mm., the temperature at 40° C. and the diffusion distance x at 0.1 cm. for all examples. Table 1 summarizes essential data for these examples. Diffusion rates for A and B calculated by Equations (21)-(24) are tabulated under method 3 in Table 1, and are considered to be correct solutions to these examples to be used for testing the approximate methods.

In general there are several possible solutions to (23) and (24), and the single function obtained by combining these equations may become discontinuous in certain regions where solutions would be expected, so that selection of the proper set of solutions is not always obvious. Greatest difficulty in this respect is encountered in cases of nearly equimolal counterdiffusion. Selection of $N_A = -N_B$ always satisfies Equations (23) and (24) so that the equimolal roots have no unique significance. For Example I three sets of solutions are obtained as listed in Table I. The set of solutions for Example I indicated as preferred was selected on the basis of agreement with rates calculated by the second approximate method of calculation, after agreement of this method

with exact solutions for the other examples had been established. In cases where diffusion rates of A and B are appreciably different in magnitude little

difficulty is encountered in the selection of reasonable solutions.

In order to obtain nonambiguous solutions for cases of nearly equimolal counterdiffusion Examples II and III were established with small partial pressure gradients among the diffusing gases. Under these conditions the right-hand members of Equations (21) and (22) are essentially constant so that these equations may be integrated directly with little error. The resulting two simultaneous linear equations are readily solved for N_A and N_B.

Method 1. In order to indicate the reasoning employed in the development of the final approximate calculation procedure it is necessary first to describe results obtained with what may be termed Method 1. This method employs basic Equations (15) and (16) with effective diffusion coefficients based on Equation (6) using a gas composition taken as the arithmetic average of the terminal conditions of the film. Thus for diffusion of A:

$$D''_A = \frac{1 - y'_A}{\frac{y'_B}{D_{AB}} + \frac{y'_C}{D_{AC}} + \frac{y'_D}{D_{AD}} + \cdots} \quad (25)$$

where:

$$y'_A = \frac{y_{A_1} + y_{A_2}}{2}, \; y'_B = \frac{y_{B_1} + y_{B_2}}{2}$$

$$\cdots, \text{ etc.}$$

Results obtained by this method for the test examples are listed under Method 1 in Table 1 where they may be compared with the correct results obtained by Method 3. Inspection of results of the two methods indicates good agreement when the component under consideration is diffusing at a relatively high rate compared with that of the other component, and poor agreement otherwise. The first possible explanation which suggested itself for this behavior was that the assumption of the arithmetic mean film composition was incorrect in defining D'' and that a more accurate evaluation might bring the results into line. Accordingly Example IV was considered in detail. In this example the diffusion of carbon dioxide is essentially that of diffusion of one gas through a second stagnant gas (hydrogen) since the partial pressure of water vapor is small. By integration of Equation (2) therefore the partial pressure of carbon dioxide could be obtained as a function of x as shown in Figure 3. Establishment of the carbon dioxide gradient permitted the calculation of D' for water vapor by Equation (6) as a function of x as also shown in Figure 3. From this plot of D' an average value D'' was determined
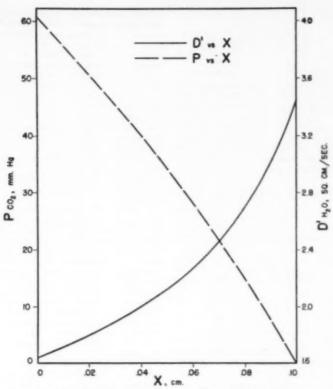
Fig. 3. Carbon Dioxide Pressure and Water Vapor Diffusivity (Eq. 6) as a Function of x for Example IV.

by graphical evaluation of the integral in the relation

$$D''_{H_2O} \approx \frac{x_2 - x_1}{\int_1^2 \dfrac{dx}{D'_{H_2O}}}$$

By this procedure D''_{H_2O} was found to be 2.15 sq.cm./sec. compared with 2.22 sq.cm./sec. calculated by Equation (25). Since this small difference in D'' could not account for the large error in the calculated rate of water diffusion it was concluded that the method as thus far developed does not completely account for the effects of simultaneous diffusion of various components.

Further insight into these counter-diffusional effects may be gained by consideration of Examples VI and VII of Table 1 in detail. In Example VI the film pressure factor for the diffusion of A as calculated under Method 1 is 288 mm. whereas the actual arithmetic average partial pressures of B and C are 75 and 72.5 mm. respectively giving a total of only 147.5 mm. Since gas C is stagnant it might be argued that its effective partial pressure in offering a resistance

to the diffusion of A is approximately its average value of 72.5 mm., and that the effective partial pressure of B is therefore 288-72.5 or 215.5 mm. Thus the effective partial pressure, or activity, of carbon dioxide in the diffusion zone with respect to water diffusion is much greater than its true average partial pressure would indicate. Therefore, if it is assumed that the average diffusion coefficient calculated by Equation (25) should be based upon the effective activities or effective partial pressures of the components in the diffusion zone rather than upon the true average composition, the effective diffusion coefficient would be lower than the value used in Method 1. This would account for the high value for the rate of diffusion of A obtained by Method 1 compared with the correct result obtained by Method 3. A similar viewpoint applied to Example VII indicates that the effective film composition for diffusion of B should be richer in hydrogen than was actually used in the diffusion coefficient calculation and that the rate calculated by Method 1 should be lower than the correct value. This conclusion is in agree-

TABLE 1.—TEST EXAMPLES * AND COMPARISON OF CALCULATION METHODS

	Gases			Partial Pressures at Point 1			Partial Pressures at Point 2			Diffusion Coefficients,† sq.cm./sec.			Calculated Diffusion Rates, (g.moles)/(sec.)(sq.cm.) × 10⁶								
													Method 1			Method 2			Method 3		
Example	A	B	C	A	B	C	A	B	C	A-B	A-C	B-C	N_A	N_B	N_A	N_B	N_A	N_B
I	CO₂	H₂O	H₂	60.44	0	89.56	0	55	95	0.9220	2.7064	3.4576	5.80	5.80	5.84		5.84	−5.64
															4.962		4.962	−4.652
															4.14		4.14	−3.93
II	CO₂	H₂O	H₂	31	49.035	49.945	49	50.945	50.035	0.9220	2.7064	3.4576	0.141	0.141	0.332	−0.106	0.331	−0.109
III	CO₂	H₂O	H₂	50.5	49.528	49.972	49.5	50.472	50.028	0.9220	2.7064	3.4576	0.0704	0.0704	0.0665	−0.0630	0.0663	−0.0529
IV	CO₂	H₂O	H₂	60.44	0	89.56	1	5	149	0.9220	2.7064	3.4576	10.58	−0.0828	10.53	−0.0431	10.52	−0.0443
V	CO₂	H₂O	H₂	60.44	0	89.56	5	5	145	0.9220	2.7064	3.4576	10.16	−0.420	10.16	−0.205	10.21	−0.225
VI	H₂O	CO₂	H₂	0	100	50	50	50	95	0.9220	3.4576	2.7064	13.06	13.06	−0.105	13.21	−0.105	13.42
VII	H₂	H₂O	CO₂	60.44	0	89.56	0	5	145	3.4576	2.7064	0.9220	10.73	−0.138	10.70	−0.243	10.68	−0.231
VIII	H₂O	CO₂	H₂	80	2	68	0	5	150	0.9220	2.7064	3.4576	15.73	0.330	15.85	0.473	16.09	0.434

* For all examples: Diffusion distance $x = x_2 - x_1 = 0.1$ cm. Total Pressure $= 150$ mm. Hg. Temperature $= 45°$ C.

† Diffusion coefficients are estimated and do not correspond to true values.

‡ Preferred Solution.

Method 1—Equations (15), (16) and (25).
Method 2—Equations (15), (16) and (20)-(28).
Method 3—Exact Solution, Equations (21)-(24).

ment with the calculations. Thus a modified effective film composition should be used for the diffusion of each component in a mixture in the calculation of the average diffusion coefficient. It should be emphasized at this point that this interpretation is offered only in a qualitative sense with no pretense of rigor and that the method of calculating the effective film composition developed here has been somewhat arbitrarily selected on the basis of the improvement shown in accuracy of the calculations on examples selected.

Method 2—General Method of Calculation. A general method of calculation using an effective film composition in evaluation of the effective diffusion coefficient will now be described. Equations will be presented in general form applicable to a system of any number of components, although it should be emphasized that the method has been developed and tested by comparison with the exact solution for three-component systems only. Diffusion of component A in a multicomponent mixture will be designated in the formulation of the equations, although the equations can be made applicable to any component by appropriate changes in subscripts.

The rate equation is given by Equation (15) of the preceding section

$$N_A = \frac{D''_A P}{RTx(p_f)_A} (p_{A_1} - p_{A_2})$$
(15)

and where $(p_f)_A$ is defined by Equation (16) and D''_A is defined by the following relations, Equations (26)-(38):

$$D''_A = \frac{1}{\dfrac{y''_B}{D_{AB}} + \dfrac{y''_C}{D_{AC}} + \dfrac{y''_D}{D_{AD}} + \dots}$$
(26)

where

$$y''_B = \frac{(p_f)_{AB}}{(p_f)_{AB} + (p_f)_{AC} + (p_f)_{AD} + \dots}$$
(27)

$$y''_C = \frac{(p_f)_{AC}}{(p_f)_{AB} + (p_f)_{AC} + (p_f)_{AD} + \dots}$$
(28)

$$y''_D = \frac{(p_f)_{AD}}{(p_f)_{AB} + (p_f)_{AC} + (p_f)_{AD} + \dots}$$
(29)

and y''_B, y''_C, y''_D ... are the effective mole fractions of the components other than A in the film, considering these components as a gas separate from A. $(p_f)_{AB}$, $(p_f)_{AC}$, $(p_f)_{AD}$ ... may be termed "partial film pressure factors" for diffusion of A into the gases designated by the subscripts. Partial film pressure factors are further defined by the relations as:

$$(p_f)_{AB} = \frac{P}{\phi_{AB}} (y'_A + y'_B)(\phi_{AB} - y'_{AB})$$
(30)

$$(p_f)_{AC} = \frac{P}{\phi_{AC}} (y'_A + y'_C)(\phi_{AC} - y'_{AC})$$
(31)

$$(p_f)_{AD} = \frac{P}{\phi_{AD}} (y'_A + y'_D)(\phi_{AD} - y'_{AD})$$
(32)

where: y'_A, y'_B, y'_C, y'_D, etc. = arithmetic mean average mole fractions of the components over the terminal conditions of the film and other quantities are defined by the following Equations (33)-(38):

$$y'_{AB} = \frac{y'_A}{y'_A + y'_B}$$
(33)

$$y'_{AC} = \frac{y'_A}{y'_A + y'_C}$$
(34)

$$y'_{AD} = \frac{y'_A}{y'_A + y'_D}$$
(35)

$$\phi_{AB} = \frac{N_A}{N_A + N_B}$$
(36)

$$\phi_{AC} = \frac{N_A}{N_A + N_C}$$
(37)

$$\phi_{AD} = \frac{N_A}{N_A + N_D}$$
(38)

The basic assumption of this method is that the effective film composition used for the calculation of D''_A is to be based on effective partial pressures of the components rather than on their true average partial pressures. These effective partial pressures are assumed equal to partial film pressure factors calculated for the diffusion of A through each component separately, as though A and the other component under consideration were present alone in the mixture, as defined by Equations (30)-(38). Effective mole fractions for each component are then defined in terms of these effective partial pressures by Equations (27)-(29), considering the components other than A as a separate gas. The effective diffusion coefficient is then defined in terms of these effective mole fractions by Equation (26), which is analogous in form to Equation (6a).

In some cases involving extremely different rates of parallel diffusion of two of the components it is possible to obtain a negative partial film pressure factor in Equations (30)-(32). In such cases it is recommended that the presence of the component giving the negative value be neglected in calculation of the effective film composition and its partial film pressure factor assumed equal to zero in Equations (27)-(29). Although this is a theoretically unsatisfactory procedure, a more suitable alternative is not apparent at this stage of the development.

Diffusion rates for the test Examples I-VIII calculated by the preceding Equations (15), (16) and (26)-(38) are listed under Method 2 in Table 1. Comparison with results of Method 3 indicates satisfactory agreement in all examples with a maximum derivation of less than 10%. Accordingly it is believed that Method 2 is adequate for most engineering applications. Although the method has been tested only with the present three-component examples, it seems reasonable to assume that equally satisfactory results would be obtained in extending the method to systems of any number of components.

It should be noted that if the diffusion coefficient did not change across the film Method 1 would be valid. For example in the problem cited by Sherwood (25) for diffusion of ammonia and water through air, where the diffusion coefficients do not vary as much with com-

TABLE 2.—COMPARISON OF CALCULATED WITH OBSERVED VISCOSITY DATA FOR SOME MULTICOMPONENT GAS MIXTURES

Components	Freon 12	So	CO₂	O₂	CO	H₂	N₂	CH₄	Higher Hydro-carbons ᵃ	Temp. °C.	Calculated Viscosity, Micropoise	Experimental Viscosity, Micropoise	% Deviation
Mole Fraction		0.333	0.333			0.333				25	195.3	185.7 (2)	+5.1
Mole Fraction	0.333		0.333				0.333			25	144.6	145.8 (3)	—0.8
Mole Fraction	0.333					0.333	0.333			25	151.0	146.9 (3)	+2.8
Mole Fraction	0.25	0.25	0.25			0.25				25	162.3	168.1 (3)	—3.4
Mole Fraction	0.25		0.25			0.25	0.25			25	148.2	147.1 (3)	+0.75
Mole Fraction			0.104		0.285	0.016	0.595			20	171.8	173.8 (13)	—1.1
Mole Fraction			0.106		0.298	0.039	0.554	0.003		20	171.5	174.3 (13)	—1.6
Mole Fraction			0.025	0.003	0.149	0.530	0.091	0.181	0.016	20	141.2	135.5 (13)	+4.2
Mole Fraction			0.048	0.003	0.264	0.172	0.442	0.036	0.005	20	169.5	171.4 (13)	—1.1

Mixture data of Herning and Zipperer were calculated on basis of pure component viscosities and binary diffusion coefficients summarized by Hirschfelder, Bird and Spotz (13). In absence of experimental data diffusion coefficients were estimated by the method of Arnold (2).
ᵃ Assumed to be ethane.

position as in the present examples, reasonably good results are obtained by Method 1.

Sample Calculation

To illustrate the use of Method 2 the essential numerical procedures in the calculation of Example VI by this method will be summarized.

Define

$$N'_A = \frac{N_A R T_{BA}}{P}, \text{ etc.}$$

A first approximation assuming $N'_B = 1.748$ and $N'_A = -0.0136$ gave $N'_B = 1.719$. The calculation for N'_A will now be shown.

Assume

$$N'_B = 1.719$$
$$N'_A = -0.0136$$
$$\phi_A = \frac{-0.0136}{1.7054} = -0.00797$$

by Equation (16)

$$(p_I)_A =$$
$$\frac{(150)(-0.00797-.03333)-(.00797-0)}{-0.00797 \ln \frac{(-0.00797-0.03333)}{-0.00797}}$$
$$= 381 \text{ mm.}$$

by Equation (30)

$$(p_I)_{AB} = \frac{(77.5)(-0.00797-0.03225)}{-0.00797}$$
$$= 391 \text{ mm.}$$

by Equation (31)

$$(p_I)_{AB} = \frac{(75)(1-0.03333)}{1} = 72.5 \text{ mm.}$$

by Equation (27)

$$y''_B = 0.844$$

by Equation (28)

$$y''_C = 0.156$$

by Equation (26)

$$D''_A = 1.04 \text{ sq.cm./sec.}$$

by Equation (15)

$$N'_A = -0.0137$$

and

$$N_A = -0.105 \times 10^{-5} \text{ g. moles/(sec.)(sq.cm.)}$$

Further trials give essentially no change in the calculated rates for *A* or *B*.

Gas Mixture Viscosities

Preceding calculations have shown that the general form of Equation (6) for effective diffusion coefficients is applicable provided the proper effective film composition is used. A further indication of the correctness of this equation has been obtained in connection with calculation of gas mixture viscosities. This work will be reviewed briefly to illustrate this application of Equation (6), and to present relations which may be useful when knowledge of mixture viscosities is required. Hirschfelder,

Bird and Spotz (*14, 15*) have recently evaluated the transport integrals of Chapman and Cowling to give alternate viscosity relations.

Buddenberg and the author (*4*) have developed the following equation for viscosities of binary mixtures.

$$\mu_M = \frac{\mu_1}{1 + 1.385 \frac{X_2}{X_1} \frac{\mu_1}{\rho_1 D_{12}}}$$
$$+ \frac{\mu_2}{1 + 1.385 \frac{X_1}{X_2} \frac{\mu_2}{\rho_1 D_{12}}} \quad (39)$$

where

μ_M = viscosity of mixture

μ_1, μ_2 = viscosity of pure components 1 and 2 at temperature and total pressure of mixture

X_1, X_2 = mole fractions of components 1 and 2

ρ_1, ρ_2 = densities of pure components 1 and 2 at temperature and *total* pressure of mixture

Extending (39) to a mixture of *n* components

$$\mu_M = \frac{\mu_1}{1 + \frac{1-X_1}{X_1}\left(\frac{1.385 \mu_1}{\rho_1 D_{1-M}}\right)}$$
$$+ \frac{\mu_2}{1 + \frac{1-X_2}{X_2}\left(\frac{1.385 \mu_2}{\rho_2 D_{2-M}}\right)}$$
$$+ \ldots \quad (40)$$

where D_{1-M}, D_{2-M}, etc., represents average diffusion coefficient of components 1, 2, etc., with respect to the total gas mixture.

Equation (6) may be written:

$$D_{1-M} = \frac{1-X_1}{\frac{X_2}{D_{12}} + \frac{X_3}{D_{13}} + \frac{X_4}{D_{14}} + \ldots} \quad (41)$$

Equations (40) and (41) may be combined to give a general equation for

gas mixture viscosities.* For three components:

$$\mu_M = \frac{\mu_1}{1 + \frac{1.385 \mu_1}{X_1 \rho_1}\left(\frac{X_2}{D_{12}} + \frac{X_3}{D_{13}} + \frac{X_4}{D_{14}} + \ldots\right)}$$
$$+ \frac{\mu_2}{1 + \frac{1.385 \mu_2}{X_2 \rho_2}\left(\frac{X_1}{D_{12}} + \frac{X_3}{D_{23}} + \frac{X_4}{D_{24}} + \ldots\right)}$$
$$+ \frac{\mu_3}{1 + \frac{1.385 \mu_3}{X_3 \rho_3}\left(\frac{X_1}{D_{31}} + \frac{X_2}{D_{32}} + \frac{X_4}{D_{34}} + \ldots\right)} \quad (42)$$

and more generally for *n* components

$$\mu_M = \sum_{i=1}^{n} \frac{\mu_i}{1 + \frac{1.385 \mu_i}{X_i \rho_i} \sum_{\substack{j=1 \\ j \neq i}}^{j=n} \frac{X_j}{D_{ij}}} \quad (43)$$

Equation (43) has been applied to viscosity data for a number of multi-component systems with satisfactory results as shown in Table 2. The agreement obtained in this application, in which the gas composition used in Equation (6) is not corrected for effects of simultaneous diffusion, suggests that viscosity behavior is such that this correction is unnecessary, and is interpreted as an indirect indication of the correctness of Equation (6) as a general form.

Equipment Capacity Coefficients. The effect of gas composition and simultaneous mass transfer of components on the commonly used equipment capacity coefficients is of practical interest. In the equivalent film theory the rate of transfer of a component is defined in

* Since the original preparation of this paper the author (*33*) has effected a simplification of Equation (43) through several approximations in the kinetic theory of diffusion to reach an alternate expression for viscosity:

$$\mu_m = \sum_{i=1}^{i=n} \frac{\mu_i}{1 + \frac{1}{X_i} \sum_{\substack{j=1 \\ j \neq i}}^{j=n} X_j \varepsilon_{ij}}$$

where:

$$\varepsilon_{ij} = \frac{\left[1+\left(\frac{\mu_i}{\mu_j}\right)^{1/2}\left(\frac{M_j}{M_i}\right)^{1/4}\right]^2}{\frac{4}{\sqrt{2}}\left[1+\frac{M_i}{M_j}\right]^{1/2}}$$

M_i and M_j are the molecular weights of components *i* and *j* and other terms are defined as for Equation (43).

This equation eliminates need of diffusivity and density data and seems comparable to Equation (43) in accuracy.

terms of the mass-transfer coefficient according to the relation:

$$N_A = (k_g)_A (p_{A1} - p_{A2}) \quad (44)$$

where

N_A = moles of component A transferred per unit time per unit area

k_{g_A} = mass-transfer coefficient for component A, moles per unit time per unit area per unit partial pressure difference

The mass-transfer coefficient is defined in terms of effective film thickness as follows:

$$(k_g)_A = \frac{D''_A P}{RT x_A (p_f)_A} \quad (45)$$

where:

D''_A = properly defined effective diffusion coefficient for component A, i.e., defined in general through Equations (26)-(38) or, under special conditions previously cited, by Equation (25)

x_A = effective film thickness for diffusion of A, and other terms are as previously defined

Similar equations are applicable to each component of the mixture.

For design calculations the mass-transfer coefficient may often be predicted from correlations based on experiments in apparatus of the type under consideration. Since these experiments usually involve the relatively simple cases of diffusion such as the stagnant gas case it is important that data correlations be put in a form such that they may be correctly applied to more complex types of diffusion. For a number of cases of mass transfer in forced convection (12, 18, 19) the effective film thickness may be correlated through dimensionless groups in equations of the general form:

$$\frac{d}{x_A} = C \left(\frac{dG}{\mu} \right)^a \left(\frac{\mu}{\rho D''_A} \right)^b_f \quad (46)$$

where

d = some characteristic length dimension for system

$\left(\dfrac{dG}{\mu} \right)$ = a characteristic Reynolds number for system

$\left(\dfrac{\mu}{\rho D''_A} \right)_f$ = Schmidt number for component A evaluated on the basis of average composition of gas film

a, b, and C are constants generally characteristic of the

type of apparatus under consideration

Under conditions of ideal gas behavior and negligible change in fluid properties over the gas film, correlations involving the Chilton-Colburn j number (5, 11, 32) as a function of Reynolds number are equivalent to use of Equation (46) with exponent b assumed to be $\frac{2}{3}$ since:

$$(j_d)_A = \frac{(k_g)_A (p_f)_A M_m}{G} \left(\frac{\mu}{\rho D''_A} \right)^{\frac{2}{3}}_f$$

$$= \frac{\left(\dfrac{d}{x_A} \right)}{\left(\dfrac{dG}{\mu} \right) \left(\dfrac{\mu}{\rho D''_A} \right)^{\frac{1}{3}}_f}$$

$$= C \left(\frac{dG}{\mu} \right)^{a-1} \quad (47)$$

where

M_m = average molecular weight of gas stream

G = mass velocity based on superficial cross-sectional area perpendicular to flow in apparatus

Recent work is believed to indicate that the assumption of exponent $b = \frac{2}{3}$ as required by the definition of the j number may not be generally applicable, and rather that the exponent is a function of shape and orientation of mass-transfer surfaces and turbulence characteristics of the flowing gas (9). Values of b in Equation (46), or equivalent exponents in other correlational forms, have been variously reported as follows: 0.44 for vaporization of liquids in a wetted wall column (12), 0.46 mass transfer from single cylinders transverse to a flowing gas (28), 0.83 for vaporization of flowing liquids in a packed column (22), and 0.9 for gas phase mass transfer in fluidized beds (23). Although these variations may be exaggerated by errors in experimental measurements, some variation in b might be anticipated in light of either the various analogies between transfer of heat, mass and momentum (7, 26) or the theory of combined eddy and molecular diffusion as related to interphase mass transfer (26, 27). It is therefore concluded that due caution should be observed in defining the exponents in Equations (46) or (47) on the basis of results in one type of system and ex-

tending them to a system of markedly different characteristics.

A general problem in equipment design is to utilize the "point value" mass-transfer coefficients just discussed in the calculation of the apparatus size required to effect a desired separation. The following development will be presented for absorption or desorption of gases in a packed tower, but the resulting equations are not so restricted in their application. Consider a differential element of height, dZ, over which transfer is occurring from a gas stream into a liquid. The liquid and gas are in countercurrent flow through the packing.

For component A

$$N_A = (k_g)_A P (y_A - y_{Ai}) \quad (48)$$

where

y_A and y_{Ai} are the mole fractions of A in main gas stream and at gas-liquid interface respectively, i.e., the boundary conditions of the film

For steady-state operation

$$(k_g)_A a_v (y_A - y_{Ai}) P dZ = -d(G_m y_A) \quad (49)$$

where

G_m = molal gas mass velocity based upon total tower cross section perpendicular to flow

a_v = effective mass-transfer surface per unit packed volume

Z = tower height

Introducing ϕ_A as previously defined and combining Equations (48) and (49)

$$dZ = \frac{-G_m dy_A}{\left(1 - \dfrac{y_A}{\phi_A} \right) (k_g)_A a_v P (y_A - y_{Ai})} \quad (50)$$

The tower height may be calculated in general by graphical integration of Equation (50), although this is not necessarily the most convenient method.

Multiplying the numerator and denominator of (50) by $(p_f)_A$ and rearranging gives

$$dZ = \left(\frac{G_m}{(k_g)_A a_v (p_f)_A} \right) \left[\frac{-(y_f)_A dy}{\left(1 - \dfrac{y_A}{\phi_A} \right) (y_A - y_{Ai})} \right] \quad (51)$$

where

$$(y_f)_A = \frac{(p_f)_A}{P}$$

The first term of the right-hand member of (51) is a general expression for the height of a transfer unit, $(H_g)_A$ (6).

If an average value of $(H_g)_A$ may be used over the tower:

$$Z = Z_2 - Z_1 = (H_g)_A \int_{y_{A2}}^{y_{A1}} \frac{(y_f)_A dy_A}{\left(1 - \frac{y_A}{\phi_A}\right)(y_A - y_{Ai})} \qquad (52)$$

and the integral in the right-hand term is the general expression for the number of gas film transfer units, $(N_g)_A$. It should be noted that in Equations (48)-(56) subscripts 1 and 2 refer to terminal conditions of the tower, and not to terminal conditions of the gas film as used in the preceding equations for pure diffusion. Compositions at the film boundaries are designated here as y_A and y_{Ai}.

For equimolal counterdiffusion

$$\phi_A = \text{ and } (p_f)_A = P$$

$$(H_g)_A = \frac{G_m}{(k_g)_A a_v P} \qquad (53)$$

$$(N_g)_A = \int_2^1 \frac{dy_A}{y_A - y_{Ai}} \qquad (54)$$

For diffusion of one gas through a second stagnant gas

$$\phi_A = 1 \text{ and } (p_f)_A = (P - p_A)_{lm}$$

$$(H_g)_A = \frac{G_m}{(k_g)_A a_v (P - p_A)_{lm}} \qquad (55)$$

$$(N_g)_A = \int_2^1 \frac{(1 - y_A)_{lm} dy_A}{(1 - y_A)(y_A - y_{Ai})} \qquad (56)$$

Equation (52) indicates that the number of transfer units is influenced by simultaneous diffusion of various components through the term y_f in the numerator and $1 - y/\phi$ in the denominator.

Written in terms of the effective film thickness the expression for the $(H_g)_A$ becomes

$$(H_g)_A = \frac{G_m R T x_A}{D''_A P a_v} \qquad (57)$$

$(H_g)_A$ is therefore independent of pressure for ideal gases and independent of the effects of simultaneous diffusion of the various components except insofar as they may affect the effective diffusion coefficient D''_A. A variation in D''_A influences x_A slightly according to Equation (46) and affects $(H_g)_A$ correspondingly.

Numerical values of (H_g) may be calculated by Equation (57) in conjunction with an equation of the form of (46) or other suitable film thickness correlation for the apparatus involved. Alternately, H_g may be expressed directly by suitable combinations of dimensionless groups, as is more frequently the case in the literature.

For many problems encountered in distillation or absorption the more simple relations (53)-(56) may be used. However, if it is necessary to consider the effects of simultaneous mass transfer of all components of a mixture the more general relations (50)-(52) are applicable. In use of these latter equations a trial-and-error solution may be necessary when values of ϕ for the various components are not known in advance.

Comment

It is recognized that the calculation procedures of this paper are not completely rigorous. Validity of the Stefan-Maxwell equations, assumed as a basis for development of the methods, has not been established experimentally in multicomponent systems for other than the stagnant gas case cited here. Further experimental studies are needed to determine the true nature of these diffusional processes. Use of the methods developed here is believed to be justified mainly on the basis of convenience and present lack of usable exact equations.

Acknowledgment

Notation *

a = coefficient in chemical reaction, or exponent on Reynolds number

a_v = effective mass-transfer area per unit packed volume, sq.cm./cu.cm.

b = coefficient in chemical reaction, or exponent on Schmidt number

C = empirical constant

d = characteristic length or diameter, cm.

D = diffusion coefficient, sq.cm./sec.

D' = point value effective diffusion coefficient with respect to a mixture

* Metric units have been used in notation to correspond to the units used in calculation of examples.

D'' = integrated average effective diffusion coefficient over diffusion path, sq.cm./sec.

G = mass velocity based on superficial tower cross section, g./(sq.cm.)(sec.)

G_m = molal mass velocity, g. mole/(sq.cm.)(sec.)

H_g = height of a transfer unit for gas film, cm.

k_g = gas-film mass-transfer coefficient, g. moles/(sec.)(sq.cm.)(atm.)

M = molecular weight

N = rate of diffusion, g. mole/(sq.cm.)(sec.)

N_g = number of gas-film transfer units

p = partial pressure, atm.

p_f = film-pressure factor, atm.

R = gas constant, (cu.cm.)(atm.)/(g.mole)(°K.)

r = coefficient in chemical reaction

s = coefficient in chemical reaction

T = temperature, °K.

U = average velocity in the x direction, sq.cm./sec.

x = distance in direction of diffusion or effective film thickness, cm.

X = mole fraction

y = mole fraction

y' = arithmetic mean mole fraction based on terminal conditions of film

y'' = effective film composition, mole fraction

Z = tower height, cm.

GREEK LETTERS:

α = proportionality constant for diffusion

μ = viscosity, g./(cm.)(sec.)

ϕ = ratio of mass-transfer rates

ρ = density, g./cu.cm.

ρ' = partial density of a component, g./cu.cm.

θ = time, sec.

SUBSCRIPTS:

A, B, C = corresponding gases or gas pairs

$1, 2, 3$ = terminal conditions of the gas film, or terminal conditions of a tower

lm = log mean average

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Discussion

B. H. Sage (California Institute of Technology, Pasadena, Calif.): It always interests me to find that the early work of Maxwell, which was in part intuitive, has been so closely confirmed by the more recent advances in kinetic theory. For example, the basic equations the author has presented for predicting the diffusion constant for the component and multicomponents can always be derived by purely kinetic considerations, utilizing the average velocity of the molecule with one or two small additions that result from the need of consideration of the other parameters that were omitted in the basic Maxwell relationships. This always seems rather hopeful, that the earlier investigators in this field were able to arrive at results which are still considered entirely applicable.

I would perhaps suggest that in the case of situations where ideal solutions are not closely approximated, that one might like to use chemical potentials and fugacity in place of partial pressure. However, in situations at the present time, limited to states where we can predict the behavior from our elementary knowledge of the kinetic theory, this is not important.

It would also seem in the latter part of the paper where the author has considered the application of his data upon the diffusion in multicomponent systems, in some circumstances where convective effects are important, that it might be worth while to go into the utilization of the over-all material balance equation in turbulent flow, and utilize the concept of eddy diffusivity in conjunction with the value of the diffusion coefficient to determine the over-all transfer. It is believed that this method will probably, in the years to come, supplement at least our present film concept theories, and the use of such concepts as film pressure factors and the over-all application of these material transfer relationships as applied to a particular point, will permit the determination of the composition of the system as a function of position. However, it will require the use of mechanical computing equipment with limited memory features. At the present time, it appears that our analytical ability has somewhat outstripped our computational ability. I hope the work during World War II, based on the development of adequate computational aids, will permit a solution of the normal nonlinear partial differential equations with sufficient ease to permit their more widespread use in industrial applications.

J. Howard Arnold: I am glad to see that Professor Wilke and his students have attacked this complicated and obscure problem of diffusion in multicomponent systems. I am doubly glad because they have been able to make use of my unsteady-state method of diffusivity measurement.

Dr. Wilke has made two rather distinct contributions in this paper. The first one is the specification of the proper type of mean value to use when you are calculating a diffusivity in a multicomponent system from the related diffusivities in the binary systems. He might have added, also, that the expression he proposes is simply a weighted harmonic mean, as opposed to the weighted arithmetic mean suggested by Hougen and Watson; the latter is badly in error in some cases.

The second contribution is to give for diffusion rates what might be considered another mean between the two extremes. On the one hand, there are the rigorously correct, but impractically complicated equations of Gilliland. On the other hand, there are equations of various sorts that are simple, but seriously in error. Dr. Wilke has proposed something intermediate, not so complicated, but nearly as accurate as that of Gilliland. Also, I think the assumption is safe that Dr. Wilke looked over various other possibilities with due diligence and selected this one as the best available. The agreement with the rigorous equation was good. The complexity was perhaps a little more than we would like, but perhaps no more than we can expect at this stage of development.

L. M. K. Boelter (University of California, Los Angeles, Calif.): I am always pleased whenever there is progress which will allow one to establish the physical properties of substances other than water and air. Such a move is one in the right direction. You all recognize, I suspect, that the considerable source of workers who were willing to spend much time determining properties has been lost. I am referring to those people who worked on the Continent of Europe, and we must now accept a considerable fraction of that burden if technology is to go forward.

(Presented at Ninth Regional Meeting, Los Angeles, Calif.)

FLASH EQUILIBRIUM CALCULATIONS

A MODIFIED RIGOROUS METHOD

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THE purpose of flash equilibrium calculations is to determine the percentage of each component that is finally in the vapor state or in the liquid state for any given set of temperature and pressure conditions.

Assuming a fair amount of experience in calculating phase equilibria, it is usually sufficient to make from three to five trials, when using the rigorous method, in order to arrive at a satisfactory result providing the determination is not too difficult.

Various approximate methods have been proposed in the last few years for the determination of phase equilibria (2, 3, 6). Several plots have also been given for simplifying the calculations involved. These determine the values for each component (1, 4, 5).

All the approximate methods require a few trials in any case. It would appear desirable therefore to restrict oneself to the rigorous method and try to develop a means of reducing the number of trials required as well as increasing the accuracy of the subsequent guesses.

Albeit the usual method of calculating flash equilibrium compositions is rigorous, it does not fix the end-point mathematically. The proposed modification accomplishes this.

Types of Problems. Types of problems encountered which require flash equilibrium calculations may be divided as follows:

1. *Adiabatic.* Initial and final heat contents must be equal.
   Flashing by sudden release of pressure is an example.
2. *Nonadiabatic.* Heat is added or removed.
   a. Vaporization as in a reboiler
   b. Condensation as in a condenser

Usual Equations. Derivation of the usual equations for flash equilibrium calculations is straightforward, based on material balances and partial pressure relationships. It can be shown that

$$l_1 = \frac{G_1}{K_1 \dfrac{V}{L} + 1} \qquad (1)$$

and

$$v_1 = \frac{G_1}{\dfrac{L}{V K_1} + 1} = \frac{G_1 K_1}{\dfrac{L}{V} + K_1} \qquad (2)$$

and

$$L = \Sigma l_n; \qquad (1a)$$
$$V = \Sigma v_n \qquad (1b)$$

Method of Calculations. The usual method of calculating for any system may be outlined as follows:

1. From conditions of the problem, determine final pressure and assume final temperature of flash or vice versa.
2. Find equilibrium constants for each component at these conditions.
3. Assume an $\frac{L}{V}$ or $\frac{V}{L}$ ratio whichever is more desirable. It is usually preferable to choose the ratio which is greater than unity.
4. Calculate the v and l for each component.
5. Find total vapor (V) by adding up all the individual v and similarly solve for L.
6. Find ratio of V to L. If this agrees with value assumed in step 3, then proceed to following step. If not, a new assumption must be made of $\frac{L}{V}$ or $\frac{V}{L}$ and steps 4 to 6 repeated until assumed and calculated values check. It is difficult to make the proper choice of subsequent trial values of the $\frac{L}{V}$ or $\frac{V}{L}$ ratio. As an aid, a plot of the assumed value as the abscissa

and the calculated value as the ordinate, may be made. A 45° line represents the locus of the final value. By plotting each trial, the proper value can be approached in four or five trials.

7. Since total heat content before flashing plus or minus the heat added or removed (zero for adiabatic conditions), and the total heat content after flashing must be equal, a heat balance will show whether the equilibrium temperature or pressure selected was the correct one. If the heat balance does not check out, a new temperature or pressure must be selected and steps 2 to 7 repeated. For a nonadiabatic system, temperature is usually fixed by conditions of the problem, thus simplifying the calculation.

There are two major difficulties in the use of rigorous methods: First, the necessity of trial and error and second, the number of trials required as well as the difficulty of determining the value to be assigned for a subsequent trial.

The basic method of calculating flash equilibrium does not lend itself to an equation which could eliminate trial and error. Such an equation for the ratio of $\frac{L}{V}$ or $\frac{V}{L}$ would be of an order equal to the number of components present in the system. As an example, if there were four components in a given system, the form of the equation would be

$$L = \Sigma l = \frac{G_1}{K_1 \dfrac{V}{L} + 1} + \frac{G_2}{K_2 \dfrac{V}{L} + 1}$$
$$+ \frac{G_3}{K_3 \dfrac{V}{L} + 1} + \frac{G_4}{K_4 \dfrac{V}{L} + 1}$$
$$(3)$$

Since

$$F = L + V$$

$$\therefore \frac{V}{L} = \frac{F-L}{L} = \frac{F - \dfrac{G_1}{K_1\dfrac{V}{L}+1} - \dfrac{G_2}{K_2\dfrac{V}{L}+1} - \dfrac{G_3}{K_3\dfrac{V}{L}+1} - \dfrac{G_4}{K_4\dfrac{V}{L}+1}}{\dfrac{G_1}{K_1\dfrac{V}{L}+1} + \dfrac{G_2}{K_2\dfrac{V}{L}+1} + \dfrac{G_3}{K_3\dfrac{V}{L}+1} + \dfrac{G_4}{K_4\dfrac{V}{L}+1}} \tag{4}$$

Where all the terms are taken under a common denominator, the vapor to liquid ratio is of the fourth power giving four answers, for one set of conditions, of which only one would be a real and positive solution.

It is because of the foregoing that several approximate methods have been proposed recently for decreasing the amount of labor involved.

Derivation of Proposed Method

By taking the basic equation and algebraically rearranging it, the author has been able to arrive at a form which shows more clearly the trend of the calculations so that it is possible to calculate the vapor-liquid ratio in fewer trials than previously and to permit the calculator to make a better second guess.

$$L = \Sigma l = \frac{G_1}{K_1\dfrac{V}{L}+1} + \frac{G_2}{K_2\dfrac{V}{L}+1}$$
$$+ \frac{G_3}{K_3\dfrac{V}{L}+1} + \ldots \tag{5}$$

$$V = \Sigma v = \frac{G_1}{\dfrac{L}{VK_1}+1} + \frac{G_2}{\dfrac{L}{VK_2}+1}$$
$$+ \frac{G_3}{\dfrac{L}{VK_3}+1} + \ldots \tag{6}$$

$$\therefore \frac{L}{V} = \frac{\dfrac{G_1}{K_1\dfrac{V}{L}+1} + \dfrac{G_2}{K_2\dfrac{V}{L}+1} + \dfrac{G_3}{K_3\dfrac{V}{L}+1} + \ldots}{\dfrac{G_1}{\dfrac{L}{VK_1}+1} + \dfrac{G_2}{\dfrac{L}{VK_2}+1} + \dfrac{G_3}{\dfrac{L}{VK_3}+1} + \ldots} \tag{7}$$

Multiplying both sides by $\dfrac{V}{L}$:

$$1 = \frac{\dfrac{G_1}{K_1\dfrac{V}{L}+1} + \dfrac{G_2}{K_2\dfrac{V}{L}+1} + \dfrac{G_3}{K_3\dfrac{V}{L}+1} + \ldots}{\dfrac{G_1\dfrac{L}{V}}{\dfrac{L}{VK_1}+1} + \dfrac{G_2\dfrac{L}{V}}{\dfrac{L}{VK_2}+1} + \dfrac{G_3\dfrac{L}{V}}{\dfrac{L}{VK_3}+1} + \ldots} \tag{8}$$

$$1 = \frac{\dfrac{G_1}{K_1\dfrac{V}{L}+1} + \dfrac{G_2}{K_2\dfrac{V}{L}+1} + \dfrac{G_3}{K_3\dfrac{V}{L}+1} + \ldots}{\dfrac{G_1K_1}{\dfrac{V}{L}K_1+1} + \dfrac{G_2K_2}{\dfrac{V}{L}K_2+1} + \dfrac{G_3K_3}{\dfrac{V}{L}K_3+1} + \ldots} \tag{9}$$

Cross-multiplying:

$$\frac{G_1}{K_1\dfrac{V}{L}+1} + \frac{G_2}{K_2\dfrac{V}{L}+1} + \frac{G_3}{K_3\dfrac{V}{L}+1} + \ldots$$
$$= \frac{G_1K_1}{\dfrac{V}{L}K_1+1} + \frac{G_2K_2}{\dfrac{V}{L}K_2+1} + \frac{G_3K_3}{\dfrac{V}{L}K_3+1} + \ldots \tag{10}$$

Combining terms, the final form of the equation is:

$$\frac{G_1(K_1-1)}{\dfrac{V}{L}K_1+1} + \frac{G_2(K_2-1)}{\dfrac{V}{L}K_2+1} + \frac{G_3(K_3-1)}{\dfrac{V}{L}K_3+1} + \ldots = 0 \tag{11}$$

Based on the $\dfrac{L}{V}$ ratio, this equation becomes

$$\frac{G_1(K_1-1)}{\dfrac{L}{V}+K_1} + \frac{G_2(K_2-1)}{\dfrac{L}{V}+K_2} + \frac{G_3(K_3-1)}{\dfrac{L}{V}+K_3} + \ldots = 0 \tag{12}$$

In order to have a vapor-liquid mix, some components must have equilibrium constants which are less than 1.0. Thus some fractions above will have negative values. The sum of the negative fractions must equal the positive. This can happen only when the vapor-liquid ratio is correct, i.e., when the value selected for the ratio is a root of the equation.

When the value of K is zero the calculation may be completed in the same manner as when K has a greater value. However, when K equals infinity

$$\frac{G(K-1)}{RK+1}$$

becomes indeterminate in this form. It can be shown that

$$\frac{G(K-1)}{RK+1} = \frac{G\left(1-\dfrac{1}{K}\right)}{R+\dfrac{1}{K}} = \frac{G}{R} \tag{13}$$

when $K = \infty$

Advantages of the Proposed Method. Sample Calculation

1. One of the important features of this proposed method is that it provides a more definite result than the usual method. The following example will serve to illustrate this point:

A vapor passes through a partial condenser. The temperature in the accumulator is 120° F. at a pressure of 360 lb./sq.in. What is the composition of each phase leaving the accumulator?

It can be shown that the proper $\dfrac{L}{V}$ ratio is somewhat less than 1.18, say 1.17 (Table 2). Calculation by the usual method is shown in Table 1. The result of 1.372 would be acceptable in a normal set of calculations. Therefore, the calculation would have been

$$\frac{1.372}{1.17} = 1.173, \ 17.3\% \text{ in error.}$$

TABLE 1.--COMMON METHOD

Comp.	G moles/hr.	K 120° F. 360 lb./sq.in.	$\dfrac{L}{V} = 1.4$ $\dfrac{L}{V} + K$	V
A	41	∞	∞	41.0
B	317	9.4	9.9	271.7
C	73	2.6	4.0	47.5
D	443	1.9	3.3	255.1
E	175	0.82	2.22	65.2
F	2,871	0.76	2.16	1,010.2
G	155	0.37	1.77	82.4
H	16	0.13	1.53	1.4
	4,091			1,724.5

$$\frac{L}{V} = \frac{2366.5}{1724.5} = 1.372$$

An agreement of this kind would be considered good, normally. However, use of the proposed method would show that this is not the end point of the calculation.

TABLE 2.--PROPOSED METHOD

Comp.	$\dfrac{G}{R}$ moles/hr.	K	$(K-1)$	$G(K-1)$	$R = 0.715$ $\dfrac{G(K-1)}{RK+1}$	$\dfrac{G(K-1)}{RK+1}$	$R = 0.85$ $RK+1$	$\dfrac{G(K-1)}{RK+1}$
A	41	∞	∞	∞		57.5*		48.2
B	317	9.4	7.4	2,345	7.0	335.0	9.14	269.0
C	73	2.6	1.6	117	2.86	40.9	3.21	36.5
D	443	1.9	0.9	399	2.36	169.0	3.62	152.0
E	175	0.82	−0.17	−29.8	1.594	−18.7	1.706	−17.5
F	2,871	0.76	−0.24	−690.0	1.543	−446.0	1.646	−419.0
G	155	0.37	−0.63	−98.0	1.265	−77.5	1.315	−74.4
H	16	0.13	−0.87	−14.0	1.093	−12.8	1.111	−12.6
	4,091				Σ positive =	602.2		525.5
					Σ negative =	−555.0		−523.3

$\ast\ \dfrac{G}{R}$

2. This method gives an automatic check of the arithmetical accuracy of all the previous calculations. After the sum of the $\dfrac{G(K-1)}{RK+1}$ values has come out to zero and, therefore, the proper $\dfrac{V}{L}$ ratio has been selected, it is necessary to divide the column of G values by the column of $RK+1$ values in order to get the $\Sigma l = L$. Subtracting from $\Sigma G = F$ gives V. Then $\dfrac{V}{L}$ gives a check. If no arithmetic errors have been made, there should be no difference except for the small inaccuracies due to the use of a slide rule.

Values for Subsequent Trials—Usual Method. After the calculations by the usual method have been completed for the first trial, an assumed value and a calculated value become available. The calculated value is the pivotal one. If the calculated value is less than the assumed, then the correct value is less than either. Similarly, if the calculated value is greater than the assumed, then the correct value is greater than either.

The numerical value to be used in a subsequent trial is difficult to determine. The usual way is to depend on a plot of $\dfrac{V}{L}$ assumed as the abscissa vs. $\dfrac{V}{L}$

calculated as the ordinate. A straight line is drawn between the known points and the intersection value with a 45° line is obtained by interpolation or extrapolation. This is then used for a subsequent trial. This may be used where two or more calculations have already been made. However, where only one set has been completed, the second trial point has only its direction fixed, as noted previously.

The author has developed a relationship which gives fair results for the estimation of this second trial. A plot of many flash equilibrium calculations has shown that the slope of a straight line connecting several trials on a 45° plot is about one-half for $\dfrac{L}{V}$ plots. From this an equation can be derived which could be used to estimate a second trial:

Let $x = \dfrac{L}{V}$ (assumed)

$y = \dfrac{L}{V}$ (calculated)

Subscript 1 and 2 are two calculated points located on the 45° plot.

$$\therefore m = \frac{y_1 - y_2}{x_1 - x_2} = 0.5$$

$$0.5x_1 - 0.5x_2 = y_1 - y_2$$

The point of intersection of this line with the 45° line (whose equation is $y = x$) is at $y_2 = x_2$

$$\therefore y_2 = \frac{y_1 - 0.5x_1}{0.5} = 2y_1 - x_1$$

$$(14)$$

Thus

$$\text{Trial } \frac{L}{V} = 2 \times \frac{L}{V} \text{ (calculated)} - \frac{L}{V} \text{ (assumed)} \qquad (15)$$

Similarly

$$\text{Trial } \frac{V}{L} = \frac{\dfrac{V}{L} \text{ (calculated)} \times \dfrac{V}{L} \text{ (assumed)}}{2\dfrac{V}{L} \text{ (assumed)} - \dfrac{V}{L} \text{ (calculated)}} \qquad (16)$$

The trial value obtained from the above relationships can be used for the next set of calculations.

Values for Subsequent Trials—Proposed Method. In order to obtain a second value of $\dfrac{V}{L}$ to be used in the subsequent calculation, an equation is proposed which is based on the following derivation:

1. Let the sum of all the positive values be

$$\Sigma \frac{G(K-1)}{RK+1} = \Sigma^+ \qquad (17)$$

2. Assume that $\Sigma k = K^+_{av}$

3. Therefore

$$\frac{\Sigma G(K-1)}{RK^+_{av}+1} = \Sigma^+ \qquad (18)$$

4. Solving,

$$K^+_{av} = \frac{\Sigma G(K-1)\text{pos.}}{R\Sigma^+} - \frac{1}{R}$$

$$= \frac{A}{R\Sigma^+} - \frac{1}{R} \qquad (19)$$

5. Similarly,

$$K^-_{av} = \frac{\Sigma G(K-1)\text{neg.}}{R\Sigma} - \frac{1}{R}$$

$$= \frac{B}{R\Sigma^-} - \frac{1}{R} \qquad (20)$$

6. Since

$$\Sigma^- = \Sigma^+$$

$$\frac{A}{RK^+_{av}+1} = \frac{-B}{RK^-_{av}+1} \qquad (21)$$

7. Therefore

$$R = \frac{-(A+B)}{AK^-_{av} + BK^+_{av}} \qquad (22)$$

Example. In the previous example a first trial value of $R = 2.0$ had been calculated. For the second trial value the method just mentioned was used:

$A = \Sigma G(K-1)$ positive $= 2860$

$B = \Sigma G(K-1)$ negative $= -831.8$

$\Sigma^+ = 254.4 \qquad \Sigma^- = -352.6$

$$K^+_{av} = \frac{A}{\Sigma + xR} - \frac{1}{R} = \frac{2860}{254.4 \times 2.0}$$

$$- \frac{1}{2} = 5.13$$

TABLE 3.

Comp.	G Prod.	Reflux	Total G	$(K-1)$	$G(K-1)$	$B=0.427$ $RK+1$	$\dfrac{G(K-1)}{RK+1}$	L
A	41	0	41.0	∞	∞		96.2*	G
B	317	38.8	355.8	7.4	2630	4.8	573.0	77.3
C	72	22.7	95.7	1.6	153	2.11	72.6	45.4
D	443	168.6	611.9	0.9	551	1.811	304.0	238.0
E	175	102.3	277.3	−0.17	−47.1	1.356	54.5	204.0
F	2871	1740.4	4611.4	−0.24	−1106.0	1.325	836.0	3480.0
G	156	117.7	273.7	−0.63	−171.9	1.158	148.3	236.0
H	16	14.8	30.8	−0.87	−26.8	1.056	25.4	29.2
	4091	2205.5	6296.5					4409.9

$* \dfrac{G}{K} = 96.2$

Σ Positive 1044.8
Σ Negative 1044.5

TABLE 4—COMPARISON

Comp.	Moles/Hr. Liquid (no reflux)	Mole Fraction	Moles/Hr. Liquid (with reflux)	Mole Fraction
A	0.	0	0	0
B	38.8	.0176	77.3	.0176
C	22.7	.0103	45.4	.0103
D	168.6	.0765	338.0	.0768
E	102.3	.0465	204.0	.0464
F	1740.4	.7891	3480.0	.7890
G	117.7	.0533	236.0	.0535
H	14.8	.0067	29.2	.0066
	2205.5	1.0000	4409.9	1.0000

This comparison shows that the amount of reflux has no effect on compositions of vapor and liquid streams providing temperature and pressure of the accumulator remain constant.

$$K_{av} = \frac{B}{\Sigma - xR} - \frac{1}{R} = \frac{-831.8}{-352.6 \times 2.0}$$

$$-\frac{1}{2} = 0.68$$

$$R = \frac{-(2860 - 831.8)}{2860 \times 0.68 + (-831.8 \times 5.13)}$$

$$= 0.87$$

The final answer was shown to be 0.855.

In the application of this method it should be noted that it loses accuracy if the amount of the component having an infinite K value is greater than 50 mole %. (Actually K always has a finite value in any problems dealt with.)

Another method which could be used when two trials have already been made is simpler but does not give as good a trial value. This is a plot of R as the abscissa as against the difference between the positive and negative values of $\dfrac{G(K-1)}{RK+1}$. A straight line is drawn between the two values obtained and is interpolated or extrapolated.

A quick method but of limited utility is to obtain a trial value by using the ratio of the positive values of $\dfrac{G(K-1)}{RK+1}$ to the negative values and multiplying by the $\dfrac{V}{L}$ used previously.

Effect of Change of Reflux Ratio. In the determination of phase equilibria, needless calculations are often made when there is a change in the reflux ratio of a fractionating column. If the

separation drum conditions remain the same i.e., vapor product, pressure and temperature as previously, then the amount of reflux condensed by the partial condenser does not affect the final composition of the liquid reflux.

This can be illustrated in Tables 3 and 4 using the previous example.

Several "Rules of Thumb"

1. Before starting any flash equilibrium calculations, the dew point or bubble point or both should be checked. Much time has been wasted trying to obtain an equilibrium which seems almost to give an answer but never does because the system is above the dew point or below the bubble point. The proposed method shows with the first set of calculations whether the system is above the dew point or below the bubble point.

2. In checking the dew point or bubble point, whenever the calculated value $(v = Kl$ or $l = \dfrac{v}{K}$ is greater than the original, there is a mixed phase present.

3. When all equilibrium constant values are greater than one, the system is above its dew point.

4. When all the equilibrium constant values are less than unity, the system is below its dew point.

5. Sometimes, it may simplify matters to use relative volatilities. Since the base component has an alpha = 1, its $\dfrac{G(a-1)}{Ra+1}$ is equal to zero. This is especially convenient where the system contains three components and the reference component selected is the middle one. Use of relative volatilities might

also be helpful where the system has high values for the equilibrium constants.

6. At the bubble point, $\Sigma G(K-1)$ is equal to zero. Below the bubble point, it has a negative value.

Notation

$A = \Sigma G(K-1)$ for positive values

$B = \Sigma G(K-1)$ for negative values

$F = \Sigma G =$ total moles in mix

$G_n =$ moles of individual component n where $n =$ any component

$K =$ equilibrium constant

$K_{av} =$ synthetic average equilibrium constant

$L =$ total moles of liquid

$l =$ moles of a given component in liquid phase

$m =$ slope

$R = \dfrac{V}{L}$

$V =$ total moles of vapor

$v =$ moles of a given component in vapor phase

$x =$ mole fraction of a given component in liquid phase

$y =$ mole fraction of a given component in vapor phase

$\Sigma l = L$

$\Sigma v = V$

$\Sigma + = \Sigma \dfrac{G(K-1)}{RK+1}$ for positive values

$\Sigma - = \Sigma \dfrac{G(K-1)}{RK+1}$ for negative values

SUBSCRIPTS:

1,2...$n =$ component number

$F =$ in feed

$av =$ average

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REACTOR PROGRAM OF ATOMIC ENERGY COMMISSION

CURRENT STATUS AND PROBLEMS

LAWRENCE R. HAFSTAD

U. S. Atomic Energy Commission

**This article summarizes the status of the development of nuclear re-
actors by the Atomic Energy Commission, and outlines a few of the
broad problems, particularly those in the field of chemical engineering.**
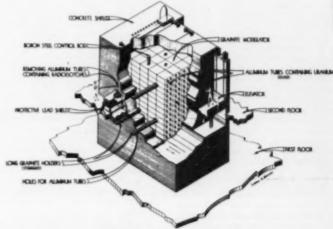
ONE problem arises from the glam-
orous nature of the subject itself.
So much has been written about atomic
—more properly nuclear—energy, it is
small wonder that misconceptions and
false hopes have arisen. For example,
it is frequently stated that the energy
in a thimbleful of water can be used to
drive the Queen Mary across the Atlan-
tic. The energy is there—no question
about that—but, in spite of recent ad-
vances in the atomic energy art, no way
is known of extracting this energy from
so small an amount of fuel.

A fundamental concept called the
"critical size" is also of importance.
The significance of this concept
lies in the fact that a certain minimum
amount of material is required before
any nuclear chain reaction can proceed.
This minimum or critical size happens
to be large, certainly large compared to
a thimble. To critical size must be added
enough shielding to make reactors safe,
and the amount of shielding required,
according to present knowledge, is of
considerable volume and weight. As a
result, machines for extracting atomic
energy, are all necessarily of relatively
large size. This, in turn, limits the
applications to those in which power is
used in correspondingly large amounts.
Thus, atomic-powered automobiles, sew-
ing machines and household gadgets

(*Presented before Atomic Energy Sym-
posium, at A.I.Ch.E. Forty-second Annual
Meeting, Pittsburgh, Pa.*)

should be promptly eliminated from
one's thinking.
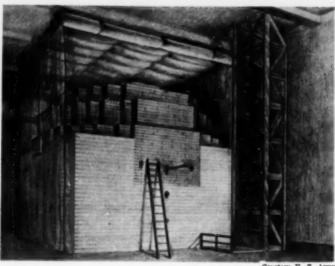
Large electric power plants supplying
our cities and atomic power reactors
would seem to be of more nearly com-
parable size. However, this does not
mean that cheap electric power from
atomic sources is assured. The money
the consumer pays for electricity goes
largely to cover costs of distribution.
Relatively, the cost of the coal or other
fuel is small. Therefore, since atomic
energy promises further reduction in
fuel costs only, it is clear that even suc-
cessful power-producing reactors could
bring little, if any, immediate reduction
in the cost of electric power to the con-
sumer.

Another problem is that of supply of
nuclear fuel. Considerable amounts of
uranium are in the earth's crust but few
concentrated deposits are yet known. A
great search is going on for them, but
up to now the world's total supply of
high-grade uranium ore, the basic
source of fuel for reactors, is definitely
limited. Also of the total uranium na-
turally occurring, only 1/140, that is,

Nuclear Reactor—Uranium "Pile"

Courtesy U. S. Army

Sketch of first self-sustaining nuclear chain reactor built under the West stands of Chicago's Stagg Field.

7/10 of 1%, is of the fissionable form, U-235. Thus, if atomic energy is to be developed as a power source, it appears that an immediate concern should be to make use of the relatively large amount of nonfissionable U-238, as well as the fissionable U-235. The possibility of doing this with a "breeder"-type reactor is an attractive one. Here emphasis is made that unless the breeding of new fuel becomes successful, the probability of civilian use of atomic power is dim. The difficulties are clearly large.

Why, then, is work going forward on the development of new and better reactors? There are several answers to this question.

The first is that, even if the future holds nothing less dismal than the continued production of atom bombs, it behooves this nation to keep in the forefront of reactor technology. At Hanford fissionable materials for atom bombs are produced in great nuclear reactors, and, sooner or later, the design of these reactors will become obsolete. Thus, considering the present international atmosphere, a large fraction of the enormous cost of the reactor program must be charged to military necessity.

A second answer to the question, why improve reactors, arises from the fact that there are times when the cheapness of power is not the controlling factor. This writer has never seen a figure for the cost per kilowatt-hour of the power

used in a B-36 airplane. Perhaps this is evidence that few persons really care. Cost is not always the dominant factor. For example, even civilian operators probably would consider secondary the cost of electric power which can be provided close to a valuable mineral deposit that is remote from other power sources.

A third answer can be given by the line of reasoning that, although nuclear reactors are now extremely costly, the newness of the technology and general unfamiliarity with this field indicate that these costs will certainly drop over the next several years. A corollary observation is that the cost of power produced from other sources appears to be following a continually rising trend. Thus, it is conceivable that sometime in the future the cost of atomic power, high though it be, will be less than the cost of power from other sources. Such facts account in part for statements by experts that the day of economic civilian atomic power is at least a decade or two away.

Fortunately, it makes little difference which of the three answers is the motivating one, the immediate actions that must be taken in reactor development are the same. Reactors now being built, or to be built soon, are all essentially experimental. They are designed to impart knowledge on their building and operation rather than for the actual use of output.

The first phase of the reactor development program is the selection and de-

sign of the first "crop" of reactors. There are four of these. These reactors and their purposes are listed here as preliminary and helpful information in studying the chemical and chemical engineering problems involved. The four reactors on which attention is being focused are:

1. *Materials Testing Reactor* which, as its name indicates, will be used in studies of materials to be employed in building reactors

2. *Ship Propulsion Reactor*, which is a land-based reactor designed as a prototype of a reactor for use in propelling submarines

3. *Experimental Breeder Reactor*, which is for use primarily to explore possibilities of breeding, that is, producing more fissionable material than is consumed. This reactor will operate with high energy, or fast, neutrons

4. *Intermediate Power Breeder Reactor*, which is designed to produce a significant amount of electric power and at the same time explore possibilities of breeding. This reactor will operate in the intermediate energy range.

This four-reactor program crystallizes the best thinking the nation has been able to muster on reactor problems since the end of World War II. However, these are in no sense the only reactors which could be built. Indeed,

many kinds of reactors are conceivable, and this fact in itself represents one of the major difficulties in formulating a rational reactor program.

It happens that the writer was associated closely with the country's guided missile work during and immediately after the war. The situation in the two fields is surprisingly similar. In the guided missile business there was a saying after the war that every time three engineers got together, a new guided missile was born. The same statement might apply to the reactor field.

Reactors can be made to operate with slow, fast, or intermediate energy neutrons; at high, low, or medium temperatures; with a half-dozen different moderators; with a half-dozen different coolants; with mechanically rigid fuel elements; or with the fuel in solution. The number of conceivable reactors is determined by the many permutations and combinations of the above-named factors.

The Commission took a step forward when it obtained the recommendation of advising scientists that four, and only four, experimental reactors should be the initial goals of the nation's program. Each reactor represents an undertaking of such difficulty that only by strong support behind each one, both in men and in money, can there be any likelihood of success. Diffusion of effort over too many reactors would lead almost certainly to disappointment and failure.

This program represents on the one hand a balance between reactors contributing to the solution of military and civilian problems and, on the other hand, a balance between reactors which use up fissionable material and reactors which promise to replenish or increase the national supply of fissionable material. It represents also a balance between a bold attempt to solve immediate problems by the engineering approach as in the Navy submarine reactor and the power-breeder, and the long-term research approach of gaining information about the behavior of materials under novel, but controllable conditions, as in the case of the materials-testing and fast neutron-breeding reactors.

The second phase of the reactor development program has been entered recently, i.e., construction. This stage will extend over a considerable period of time. Fabrication of components and auxiliaries and the erection of structures for a reactor are complex and time-consuming. The program is going forward.
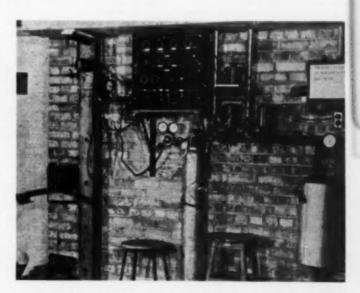
The Experimental Breeder Reactor is

Courtesy U. S. Army

Start of a neutron bombardment for producing radioisotopes in Clinton chain-reacting uranium "pile." Dr. Waldo Cohn demonstrates how samples of materials are inserted into the pile to make them radioactive.

farthest along. Nuclear design by the Argonne National Laboratory has been completed, and architect-engineering design work by the Austin Co., of Cleveland, Ohio, is more than 90% complete. The Bechtel Corp., San Francisco, Calif., has been selected as the construction contractor. Installation of utilities,

A view of the "rabbit" (pipe extending through wall at left) which is a device whereby materials can be introduced quickly and easily into the pile for neutron irradiation.

excavation work and construction of an access road are already underway for this reactor at the nation's new Reactor Testing Station in Idaho.

The Materials Testing Reactor, which is also to be located at the Reactor Testing Station, is following close behind. Scientific design has been completed jointly by the Argonne National Laboratory and the Oak Ridge National Laboratory. Detailed engineering design is being done currently in Pittsburgh by the chemical plants division of Blaw-Knox. Construction is expected to begin in the spring.

For the Ship Propulsion Reactor, engineering and development work is being carried out at Argonne and at the Westinghouse atomic power division on the site of the old Bettis Airport near Pittsburgh and detailed engineering design is scheduled to begin in about a year. Construction should be underway by 1952. The reactor will be built at the Reactor Testing Station.

As to the Intermediate Power Breeder Reactor, engineering design work has been pushed to an advanced stage by the General Electric Co. The reactor is planned for construction at the West Milton (N.Y.) site near Schenectady. Preliminary site work, now underway, is expected to be completed in time for construction of major facilities to begin early in the 1950 construction season.

To return to the question of "breeding," it is noted that of the four reactors, two are breeders and two are nonbreeders. The nonbreeders are pointed essentially toward power production. The breeder reactors represent a more subtle concept and one which, this writer believes, will be of exceptional interest to a professional group in chemical engineering.

The breeding process consists of production of fissionable material in a reactor in greater amount than is used up in the fission process which keeps the reactor operating. The hope is for the breeding of substantially *more* fissionable material than is consumed as fuel; hence the label "Operation Bootstrap." In spite of implications, breeding is not akin to perpetual motion, though it is almost as intriguing. The promise is that by feeding a reactor the relatively more abundant but nonfissionable U-238 it may be possible to keep the reactor almost indefinitely supplied with fuel. Yet there is no magic or mystery involved; the success of the breeding process depends directly on the extent to which neutron losses can be reduced.

Uranium happens to come in nature in two forms, differing only in atomic weight; namely, U-238 and U-235. Only the U-235 is fissionable. By the fission of U-235 in a reactor containing both

U-235 and U-238, excess neutrons released in the fission of U-235 may be captured by the U-238 and, as a result of the capture, convert U-238 into the fissionable new element, plutonium.

This is the process of the Hanford reactors. The important point is that less plutonium is formed than there was U-235 in the first place. In other words, in the Hanford reactors, there is a net loss of fissionable material. Only part of the energy originally available in the U-235 ends up in plutonium which can be used for bombs or other purposes; the rest of the energy is wasted.

In each fission process an average of between two and three neutrons are emitted. One of these neutrons is required to replace the original neutron and maintain the chain reaction. Another is captured by U-238 with the resultant creation of a plutonium atom. Then, there is still on the average a fraction of a neutron available to take care of leakage and losses. Such a reactor would be an ideal "converter" because, at the end of the experiment, there would be one atom of plutonium created for every atom of U-235 destroyed. However, since allowance is made for a fraction of a neutron per fission for losses, if these losses could be reduced still further, the operation would end up with more than one fissionable plutonium atom for each fissionable U-235 atom. That would put "profit" into the operation.

Continual conversion of U-235 to plutonium ultimately produces enough plutonium to build a new reactor with this element as a fuel. Now, the real breeding process can start. Beginning with a given amount of fissionable plutonium, if more plutonium is produced than is used up, the longer the operation continues the more plutonium will be found in the reactor, as long as—and this is the important point—there is an adequate reserve supply of the nonfissionable U-238 available. This is certainly not perpetual motion. This is merely a process of converting unusable U-238 into usable fissionable plutonium. However, this is no mean achievement, for the supply of U-238 is 140 times as great as that of U-235. The rate of augmentation of supply promises to be very slow at best, but the potentially available supply of fissionable material is being increased—the potential atom bomb stockpile, if you like—by a factor of 140. This is a goal worth shooting at.

Difficulties are not over, however, if in the reactor itself the breeding process is successful. There remains the problem of reprocessing the fuel elements themselves. Fission products, radioactive elements in the neighborhood of barium in the periodic table, are formed in the fission process. These are

the ashes of the reaction which, in sufficient accumulation, tend to smother the fire. Every so often the fuel elements must be removed from the reactor, chemically purified to remove fission products, refabricated and reinserted into the reactor. Losses in this chemical processing are just as effective in reducing "profit" as the losses of the neutrons themselves. The key question, therefore, is: Can the total losses, both of neutrons and in chemical processing, be kept sufficiently low so that in a complete cycle there is a net gain?

Here is where the chemical engineer comes in, and here, one obtains a glimpse of the nature of a new front opening up in the campaign for reactor development. The problem of fuel reprocessing is an enormous one, and it is almost entirely a chemical engineering problem. The cost of this processing is so high that, even if all other costs in the reactor business were kept low, it alone might keep power derived from nuclear reactors from competing economically with other fuels. It pains this writer as an engineer to see carefully fabricated and machined parts, products of many man-hours of highly skilled work, casually dissolved in acid to start the chemical purification. This procedure is even more painful to chemical engineers, so they have been suggesting reactor designs in which the laborious fuel element fabrication step can be eliminated either by placing the fuel in a solution or by utilizing techniques of transporting fluidized solids.

From such thinking there emerges the possibility of a new kind of reactor called the homogeneous reactor. All reactors constructed so far, except for one small experimental unit, have been built upon the principle of embedding fuel elements in other materials used for cooling, reflecting, and moderating. The question has been studied and restudied from time to time to see whether it would be feasible and practical to make a reactor in which these constituents are mixed together uniformly; hence the term "homogeneous reactor."

The homogeneous reactor, in its various conceivable forms, opens up entirely new vistas and areas of development in the field of reactor design. It is this writer's belief that with experience from the first four reactors under our belts the position would be favorable to attack this difficult new design and this new type will constitute a large part of the next crop of reactors. Obviously, chemical engineering must contribute heavily. It must come to grips with the fuel reprocessing problems involved in the operation of the first four reactors, and begin to take advantage of the opportunities which exist in the "homogeneous" design.

• NEWS SECTION •

SOUTHWEST IS CALLING!

THE regional meeting of the American Institute of Chemical Engineers will be held in Houston, Tex., Feb. 26-28, and March 1, 1950. Persons attending will want to visit some of the points of interest in addition to attending the technical sessions. Even though the city has amazed the industrial world with its phenomenal growth, it has managed to maintain its cultural status also.

On Jan. 1, 1950, the population of Houston was estimated at 620,000; Greater Houston at 700,000, and Harris County at 780,000. Thus it is the largest city in the South.

First, Houston is proud of the Houston Ship Channel, Turning Basin, and wharf facilities. The channel is 58 miles in length and extends from the entrance to Galveston Bay to Houston. Of this distance, about 50 miles represents the ship channel proper and 8 miles, the distance from the Bay to the Bolivar Roads entrance and the deep water of the Gulf. About half the channel proper was dredged through Galveston Bay; the re-

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HOUSTON

Thirteenth Regional Meeting

Rice Hotel
February 26-28, March 1, 1950

Article by
W. H. Tonn, Jr.

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mainder is the enlarged Buffalo Bayou. Channel has an average depth of 34 feet, and a width which allows ocean-going vessels complete maneuverability.

The banks of the channel are lined with chemical plants, oil refineries, cot-

ton compresses, and general shipside warehouses. Berthing space can accommodate 80 ocean-going vessels.

Seventeen railroads serve the city and connect with the port facilities. Tonnage handled by Port Houston last year placed it second only to New York in this country.

Near the location where the San Jacinto River empties into Buffalo Bayou (Ship Channel) and on the opposite side of the Bayou is situated the San Jacinto Battleground, Memorial Shaft, and the Battleship "Texas." It was on this spot that General Sam Houston won the battle for the independence of Texas from Mexico by defeating the Mexican General Santa Anna. The monument is the tallest single shaft in the world, towering some 570 feet into the sky. An observation room at the top and elevator provide for easy access. At the same location is the Battleship "Texas" which is open to the public and contains much of its original equipment and trophies.

Main Street, Houston's principal business thoroughfare.   Houston's new $1,750,000 City Hall

W. A. Cunningham, Chairman, Regional Meeting

As for educational facilities, Houston has its fair share. There is Rice Institute, one of the most advanced educational institutions in the nation, and its beautifully landscaped campus and fine architecture. The rapidly growing University of Houston is located in a splendid pine-wooded tract of land and has a large expansion program underway. The University of Texas School of Dentistry, Baylor College of Medicine, and Texas State University for Negroes complete the résumé.

The Texas Medical Center is still under construction, although a portion, such as Baylor College of Medicine, has been completed. This is a 163-acre site not far from the center of town. Included in it is the $12,000,000 Veterans Hospital, Hermann Hospital, the M. D. Anderson Cancer Research Laboratory, and Hermann Professional Building. Also, the University of Texas School of Dentistry is scheduled to construct a building at the site.

The gigantic Sam Houston Coliseum, which houses the Annual Houston Fat Stock Show and other events; the $2,000,000 City Hall; the Public Library

Courtesy of Houston Chamber of Commerce

Statue of General Sam Houston, commemorating the battle at San Jacinto, which stands at the entrance to Hermann Park in Houston.

with its 306,000 volumes; and Music Hall, are located within the shadows of the downtown skyscrapers.

For individuals interested in art and music, it should be remembered that the Museum of Fine Arts has an exhibition of numerous works of art, paintings, and sculpture.

Houston has 62 municipal parks and playgrounds, including four municipal golf courses, bridle paths, tennis courts, and other recreational facilities. Hermann Park is the home of a fine zoo and a museum of natural history. Many beautiful country clubs dot the city.

Within a short distance are numerous cattle ranches, oil fields, rice farms, and sulfur wells. These can be easily seen by auto trip, along with the Washburn Tunnel now under construction. This tunnel will give passage to motor vehicles between both sides of the channel at Pasadena.

Another feature of the program will be the students section. Chemical engineering students in the various Southwest colleges have been invited to present papers on various chemical engineering processes. These papers will be screened first at the colleges and those selected will be presented at the regional meeting, March 1, in competition for more than $100 offered in prizes. Included in this student section of the activities will be a luncheon in which recognized outstanding men enter into informal talks and conversation with the students.

Courtesy of Houston Chamber of Commerce

New express highway now under construction from downtown Houston to Galveston.

WHICH IS YOUR NEED...

WATER PURIFICATION
Through Distillation

- Potable Water from Sea or Brackish Water
- Pyrogen Free Water from any source
- Bacteriologically Sterile Water

CONCENTRATION FOR RECOVERY OR DISPOSAL
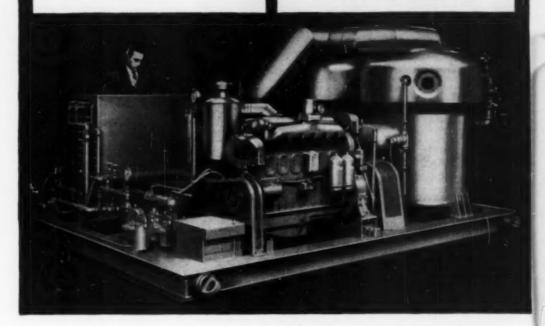
- **Concentration for Recovery**

Tanning Liquors	Penicillin Slurry
Brine	Industrial Chemicals
Amino Acids	Black Liquor from
Other solutions and substances	Pulp-Paper Mills

- **Concentration for Disposal**

Brine	Penicillin
Other solutions and substances	

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CLEAVER-BROOKS COMPRESSION STILLS solve many concentration problems. They provide economical recovery of valuable solids which are normally lost in waste

disposal. Cleaver-Brooks Stills have concentrated Tanning Liquors, Brine, Penicillin Slurry, Black Liquors from pulp mills and many other solutions.

Available in standard size units from 75 gph to 2100 gph, electric or Diesel drive. Larger sizes to fit specific applications can be constructed for field erection. Your water or concentration problem will be thoroughly studied by Cleaver-Brooks engineers and recommendations made for improving your plant production.

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BUILDERS OF EQUIPMENT FOR THE GENERATION AND UTILIZATION OF HEAT

NEW RESEARCH LABORATORY, FOOTE MINERAL COMPANY

The new three-story building of the Foote Mineral Co. on the Lincoln Highway in Berwyn, near Paoli, Pa., has been equipped to carry out basic research, product development, and engineering studies related to improvement in equipment design and process control. The top floor is devoted to chemical investigations, including analytical, inorganic, physical, and organic chemistry laboratories and glass-blowing facilities. Further development work on lithium compounds, zirconium, is the subject of major projects. The expanded facilities were shown for the first time at an informal open house held recently at the new laboratories.

COMMITTEE APPOINTED
FOR A.E.C. STATION

Appointment of a three-man Engineering Advisory Committee to assist the Idaho Operations office, U. S. Atomic Energy Commission, on the planning and development of the Reactor Testing Station was announced recently by L. E. Johnston, manager of the Idaho office.

Members of the Committee are: General L. J. Sverdrup, Chairman; Dr. H. M. Crothers; and W. W. Horner. All three have long records in the field of engineering.

General Sverdrup is president of Sverdrup and Parcel, Inc., consulting engineers, of St. Louis, Mo. During World War II he served as acting chief engineer to General MacArthur during the Leyte and Luzon invasion in the Philippines and later as Commanding General of the Engineer Construction Command in the Southwest Pacific.

Dr. Crothers is vice-president and dean of engineering, South Dakota State College, Brookings, S. D. During the war he served as deputy director of Engineering Science and Management War Training for the U. S. Government.

Mr. Horner is a partner in the firm of Horner and Shifrin, consulting engineers of St. Louis, Mo. He has been a lecturer and a professor at Washington University.

Commenting on establishment of the new committee, Mr. Johnston said:

"In the development of the Reactor Testing Station, we recognize the need for the most competent technical advice that can be obtained. Through the Engineering Advisory Committee, we are getting men with background and experience who will help us to build soundly, efficiently, and economically."

A.C.S. CONTINUATION
COURSES ANNOUNCED

The chemical education committee of the Philadelphia section of A.C.S. will present two continuation courses during the ten weeks commencing Feb. 13. Both courses will be given at the Philadelphia College of Pharmacy and Science, 43rd and Kingsessing Ave., Philadelphia.

Course 1—Chemical Engineering Thermodynamics is intended to supplement undergraduate courses in chemical thermodynamics. Eight specialists will discuss various aspects of the subject. Monday evenings—7:30-9:30.

Course 2—Instrumental Electronics is a new course organized to meet the desires of chemists who use electronic apparatus in the laboratory and who are interested in acquiring some familiarity with its mechanism of operation. Tuesday evenings—7:30-9:30.

Each course will be $10 for members of the A.C.S. and $12 per course for nonmembers.

DEADLINE—MARCH 15
ENGINEERING SURVEY

The closing date for the nation-wide survey of selected engineering personnel now being sponsored by the Engineers Joint Council for the U. S. Office of Naval Research has been announced as March 15, 1950. Up to Jan. 20 more than 60% of the 115,000 questionnaires, sent to full members of 18 national engineering societies, were answered. Following processing by the A.S.M.E., contracting agent under an O.N.R. agreement, the questionnaires will be kept in Washington as a source file of the nation's key engineers and scientists.

The survey was initiated as the result of a conference held in Washington late in 1948 attended by E.J.C. representatives and those of many other engineering agencies, at which the need was discussed for a list of key engineers working in research, development, and other scientific projects, who could be called in on a full-time or part-time basis to work on the broad scientific programs of the National Military Establishment.

A national asset, the body of facts gathered by the survey will be available to Government agencies, industrial, educational and professional society planning groups, and for other legitimate purposes.

The survey is managed by the E.J.C. Committee. C. E. Davies, secretary of A.S.M.E. is chairman.

NATIONAL CHEMICAL
EXPOSITION OF A.C.S.

The Sixth National Chemical Exposition by the Chicago section of the American Chemical Society, will be held in the Chicago Coliseum Sept. 5-9, 1950, according to an announcement by Dr. Marvin C. Rogers, research director of R. R. Donnelley and Sons Co., who is chairman of the committee in charge.

Dr. Rogers has also announced the names of the members of the Exposition Committee who will assist him in planning and conducting the chemical show. These are: Dr. George G. Lamb, Technological Institute of Northwestern University; Orville G. Linnell, R. R. Donnelley and Sons; Dr. T. U. Marron, A. B. Dick Co.; Dr. C. S. Miner, Jr., The Miner Laboratories; A. E. Schneider, The Armour Laboratories; Dr. H. W. Schultz, Swift & Co.; B. N. Rockwood, Swift & Co., and H. F. Schwarz, Sherwin-Williams Co. Details of the Exposition will be handled by the manager, James J. Doheny, newly appointed manager of the Chicago section of the A.C.S. Since 1946, Doheny has been in charge of arrangements for national meetings of the A.C.S.
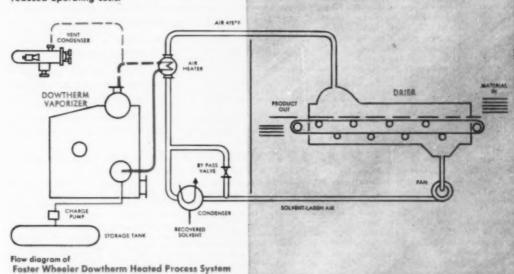
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NOYES EDITOR
J. AM. CHEM. SOC.

Dr. W. Albert Noyes, Jr., chairman, department of chemistry, University of Rochester, became editor of the Journal of the American Chemical Society on Jan. 1. He succeeds Dr. Arthur B. Lamb of Harvard University, who is retiring from the post after a 31-year service. Both Dr. Lamb and Dr. Noyes are former presidents of the society; Dr. Lamb held that office in 1933 and Dr. Noyes in 1947. Dr. Lamb will continue to serve the monthly as consulting editor. The editorial headquarters will be moved from Harvard to the University of Rochester.

STANDARD STEEL BUYS
HERSEY DRYING DIV.

Standard Steel Corp., Los Angeles, Calif., recently announced the outright purchase of the drying machinery division of Hersey Manufacturing Co., Boston, Mass. Standard Steel has for many years manufactured drying and dehydration equipment for chemicals, fertilizers and by-products.

Plans for 1950 call for manufacturing in the east as well as in the main plant at Los Angeles, and expansion of sales, service and plant facilities is already under way. H. W. Harrigan, general manager of the Hersey division, will become eastern manager of Standard Steel Corp. J. A. Boyd, chief engineer for Hersey with 25 years' experience in the design of dryers, is moving to Los Angeles to assume a similar position at Standard Steel, assisting Russell J. Love, vice-president in charge of engineering.

SLOAN FELLOWSHIP
GRANTED BY M.I.T.

Fourteen young business executives, largest group in the 13-year history of the executive development program of the Massachusetts Institute of Technology, Cambridge, Mass., will be awarded Sloan Fellowships in 1950, it was recently announced. These fellowships, awarded to outstanding young executives in a nationwide competition, cover a year of advanced study in economics and business administration at the Institute and are aimed at preparing men for higher executive responsibility.

The fellows will be drawn from both large and small companies in various types of industry throughout the country. The winners of the fellowships will participate in a special program consisting of seminars, classes, and field investigations aimed not only at increas-

ing technical managerial skill, but also at deepening their understanding of the social and economic implications of their work. They will study in the Institute's departments of business and engineering administration and economics and social science, and will have an opportunity to conduct a special investigation in some selected field of industrial activity such as finance, production, distribution, or labor relations.

Applications for the awards must be made by Feb. 24. Recipients will be selected on a competitive basis by M.I.T. Awards range up to $2700 for single men and up to $3700 for those who are married. Fellows will be in residence at the Institute in Cambridge from June 9, 1950, to June 10, 1951. Applicants must be between the ages of 30 and 35 and have at least five years' industrial experience, part of it in an executive capacity.

INDUSTRY GROUP TO
STUDY A.E.C. FILES

A working party of representatives of technical and engineering societies and business press has been selected by the Atomic Energy Commission to begin the examination of declassifiable technological information in the field of metallurgy with a view to determining its possible value to American industry. Members of the working party who have been given complete clearance to enter restricted areas and to examine restricted information are:

Keith Henney, consulting editor, *Nucleonics and Electronics* and representative, I.R.E.

S. A. Tucker, A.S.M.E.

F. J. Van Antwerpen, Editor, *Chemical Engineering Progress*, A.I.Ch.E.

Sidney Kirkpatrick, vice-president, McGraw-Hill Book Co., editor, *Chemical Engineering*

E. E. Thum, A.S.M., editor, *Metals Progress*

Establishment of the working group is the second step in a trial program set up by A.E.C. in response to the recommendation of its Industrial Advisory Group that information still classified, but potentially declassifiable and of interest to industry should be surveyed and declassified. The first step was taken with the appointment of a temporary advisory committee which met in September, 1949, with officials of the Atomic Energy Commission.

The first task before the working group will be to examine the abstract file maintained by the A.E.C. Patent Branch.

PROJECTS OF ARMOUR
RESEARCH NUMBER 296

Accomplishments in a wide variety of scientific fields make up an annual report issued recently by Armour Research Foundation of Illinois Institute of Technology, Chicago, Ill.

In the 44-page report are descriptions of 296 research projects undertaken by 630 staff members during 1948-49. Sponsored by industry, government agencies, and the Foundation itself, the projects represent an investment of more than $4,000,000. This was an increase of 22 per cent over the previous year.

A monthly analysis of dust fall for the city of Chicago, sponsorship of a state-wide conference of industrial executives to discuss research, and tracing hundreds of chemicals for scientists the world over by the National Registry of Rare Chemicals comprised some of the Foundation activities.

The Foundation is a nonendowed, nonprofit scientific research and development organization founded in 1936 to serve industry. Any excess of income over expenses is used to finance fundamental research in the public interest and to pay for the Foundation's continuing public service activities. Dr. Haldon A. Leedy, Foundation director, announced a budgeted expansion of 15 per cent for the 1949-50 fiscal year.

EQUILIBRIUM CHARTS
NOW AVAILABLE

To facilitate application of the methods for calculating vapor-liquid equilibria at high pressures presented by Smith and Watson, *Chemical Engineering Progress*, **45**, 494 (1949) the Sinclair Refining Co. has prepared a set of 21 Equilibrium Constant Charts for the paraffin hydrocarbons. These charts cover the boiling point range from methane to 700° F, pressures up to 10,000 lb./sq.in., and temperatures from 0° to 1000° F.

A short paper by K. A. Smith and R. B. Smith discussing use of these charts appeared in the December issue of *Petroleum Processing*. A reprint booklet is now available which includes a reprint of this paper together with 41 full-page working charts comprising the Sinclair Equilibrium Constant Charts and large scale reproductions of the fugacity and activity coefficient charts of the original paper. Copies of this booklet may be ordered from *Petroleum Processing*, 1213 West Third Street, Cleveland 13, Ohio, at a cost of $1.50 each, postpaid.

ENGINEERS JOINT COUNCIL REPORTS

FOR the first time since its organization, Engineers Joint Council has issued an annual report. It was compiled in the office of the 1949 Secretary for EJC, reviewed by the Council's Committee of Secretaries and received for the record by Engineers Joint Council at its meeting Jan. 20, 1950. An extract follows:

Engineers Joint Council is an evolution from the one-time Joint Conference Committee. The latter, established in October, 1941, to promote cooperation among the four Founder Societies, was composed of the Presidents and the Secretaries of the Societies. In 1942, the American Institute of Chemical Engineers became the fifth participating organization. As the scope of activity expanded, stimulated by the urgent need for a united front of the profession during the years of World War II, the "Conference" concluded that its function was of greater significance than the name implied and, in September, 1945, Engineers Joint Council came into being. Thus, although Council had its beginning in 1941, it has just rounded out its fourth full year under its present identity. They have been years of increasingly vigorous and constructive activity. At the year's end the total membership of the five Societies of EJC was 111,000.

A major internal step in the development of Council during the past year has been ratification by the governing bodies of the constituent societies of a formal constitution to replace the previous rather informal by-laws. Under its terms, Council now is composed of the two most available recent past-presidents and the secretary of each of the five principal national engineering Societies. Previously, the current President of the Societies had served with one Past-President, with the result that these Presidents carried an unduly heavy load during their first year on Council. The new arrangement is a definite improvement.

The officers and members of Engineers Joint Council for 1949 were as listed below:

President—R. E. Dougherty
Secretary—W. N. Carey
Vice-President—Scott Turner
Treasurer—E. L. Chandler

American Society of Civil Engineers

R. E. Dougherty, Past-President
E. M. Hastings, Past-President

W. N. Carey, Executive Secretary
C. S. Proctor, Alternate

American Institute of Mining & Metallurgical Engineers

H. G. Moulton, Past-President
Scott Turner, Past-President
E. H. Robie, Secretary
J. L. Head, Alternate

The American Society of Mechanical Engineers

E. G. Bailey, Past-President
E. W. O'Brien, Past-President
C. E. Davies, Secretary
R. M. Gates, Alternate

American Institute of Electrical Engineers

J. E. Housley, Past-President
D. B. Hull, Past-President
H. H. Henline, Secretary
J. F. Fairman, Alternate

American Institute of Chemical Engineers

S. D. Kirkpatrick, Past-President
L. W. Bass, Past-President
S. L. Tyler, Secretary
D. B. Keyes, Alternate

In former years, meetings were held at irregular intervals, depending upon the amount and importance of work confronting Council. Because of the volume of matters requiring attention, and better to systematize procedure, it became increasingly evident that a regular schedule for meetings should be established. Consequently, the third Fridays of odd months now are official dates for EJC meetings. Meetings customarily are held in the Engineering Societies' Building in New York.

The Constitution (Article I, Section 2) states that the objectives of the Council shall be:

a. To advance the general welfare of mankind through the available resources and creative ability of the engineering profession

b. To promote cooperation among the various branches of the engineering profession

c. To develop sound public policies respecting national and international affairs, wherein the engineering profession can be helpful through the services of the members of the engineering profession

Upon that foundation, Engineers Joint Council, in its role as the cooperative, fed-erating agency for the national societies representing the five basic branches of the profession, has been active in a wide variety of matters of national and international scope. Criteria for determining participation in any such matter are: (1) that it shall be of wide interest to the profession, and (2) that it shall be a matter having a direct impact on the profession or one regarding which it is believed that action by the profession can contribute to the public interest.

For years, EJC has been devoting study to stimulation of greater solidarity of the engineering profession. One of the most significant steps taken during 1949 was the calling of a conference, under the general auspices of Council's Committee on Unity of the Engineering Profession, for the purpose of developing specific recommendations as to measures for achieving further unity. Sixteen of the principal national engineering Societies were invited to participate. At the meeting, Oct. 20, 1949, fourteen Societies were represented, delegates of the other two being unavoidably absent. As the result of general discussion, a committee was appointed to study the relative advantages of several suggested courses of action and to report back to the conference at a meeting to be held early in 1950.

The small expense incurred in connection with the work of EJC is significant. Each year a budget is established and the necessary funds are prorated among the constituent societies in proportion to dues collections by each for the preceding fiscal year. While expenditures may vary within rather wide limits, depending on the character of projects undertaken, it is interesting to note that for 1948 and 1949, the total each year was $3900, in round numbers.

Disbursements for 1949 were broken down as follows:

General operating expense and
secretarial service $2,498.28
General survey committee ... 697.23
Committee on collective bar-
gaining 250.00
Labor legislation panel 376.49
Science legislation panel 73.52
Committee on international
relations 33.88

Total $3,929.40

That so much can be done with so little financial outlay, is a tribute to the large number of men who have contributed . . . their time and abilities in carrying out the programs of Council. A review of the list of personnel will indicate the number and caliber of men who are devoting themselves unselfishly to the objectives of the Council and thus rendering service to their profession, to their country, and to the world. The full extent of that service cannot be appreciated except by those who have been closely in contact with the work.

A copy of the complete report may be had on request from the Secretary of any one of the five EJC Societies, or from the Secretary of EJC for 1950, E. H. Robie, c/o American Institute of Mining & Metallurgical Engineers (AIME), 33 West 39th Street, New York 18, New York.

E. J. C. COMMITTEES

Constitution and By-Laws
General Survey
Increased Unity in the Engineering Profession
Labor Legislation
Science Legislation
Engineers Cooperating in Medical Research
Engineers in Civil Service

Selective Service
Federal Income Tax
National Water Policy
Fuel Resources
National Engineers
International Relations
UNESCO

Consultative Status with the United Nations

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MARGINAL NOTES

Present Status of Chemicals from Petroleum

The Petroleum Chemicals Industry. Richard Frank Goldstein; E. & F. N. Spon, Ltd., 57 Haymarket, SW 1, London. To be published in the United States by John Wiley & Son, New York 16, N. Y.

Reviewed by T. Q. Eliot, Research Department, Stanolind Oil and Gas Company, Tulsa, Okla.

DR. GOLDSTEIN has attempted, in this monograph, to give a reasonably complete status of the art of manufacturing chemicals from petroleum, with special emphasis on the developments during the last ten years. The bulk of the manuscript was written in 1946, and additions have been made to bring the subject matter up to date by the end of 1947.

While the main objective of the monograph has been to survey the fields of industrial organic chemistry in which hydrocarbons are the preferred starting material, consideration is given also to processes using competitive raw materials (e.g., ethyl alcohol by fermentation). Since commercial practice has led chiefly to aliphatics, emphasis is placed on this division of organic chemistry. However, one chapter is devoted to the manufacture of aromatics.

The volume comprises 20 chapters devoted to sources of petroleum hydrocarbons, industrially important reactions such as oxidation, chlorination, nitration, etc., as applied to both paraffins and olefins, and the manufacture and reactions of naphthenes and aromatics. Chemical classes including aldehydes, ketones, acids and anhydrides, olefins oxides, nitriles and amines are discussed in separate chapters. A brief review of synthesis and reactions of $CO—H_2$ mixtures is included with primary emphasis on the German art. Appendices on routes to petroleum chemicals and statistics of production and consumption are included, as well as a number of flow charts covering principal reactions of particular materials such as acetone, ethylene oxide, etc.

Dr. Goldstein has attempted a major task, with some degree of success. The text is well organized, considering the complexity of the subject. The small number of references and patents is disappointing, especially since many of the latter are British. In fact, it is somewhat surprising to find a British monograph on this subject, for the art of manufacturing chemicals from petroleum was developed largely in the United States. It is probably this situation which has resulted in a book which is largely a review of the German and U. S. literature during 1935-47. In many cases, a more accurate picture of an individual development may be obtained from a survey of "Chemical Abstracts." The scope of the subject matter precludes a detailed discussion of any subject, and, in general, the text indicates a benign acceptance of everything in the literature; no attempt is made to sift, criticize or straighten out the facts. The volume should be useful, however, as an auxiliary reference to Carlton Ellis's "Chemistry of Petroleum Derivatives."

Clay and Its Applications

Kaolin Clays and Their Industrial Uses. J. M. Huber Corp., New York (1949). 141 pp.

Reviewed by John B. Calkin, Director, Department of Industrial Cooperation, University of Maine, Orono, Me.

THE opening words of this book stating that "this book is an example of the widespread yet integrated nature of the Huber Operations," keynotes this work from "The Huber Laboratories in Texas, Georgia, and New York." The book is distributed free.

The book is divided into six parts covering (1) definitions, occurrence, composition and production of clay; and uses of clay in (2) rubber (3) paper (4) adhesives (5) insecticides and (6) ceramics. Greatest emphasis is placed on rubber and paper with further information on the other uses to follow in another volume. Forty-five (45) literature references are cited with some errors occurring in these such as TAPPI Method T 459 m-42 (indicates 1942 tentative) which was revised in 1945 and again in 1948.

The work gives a good start in discussing the subject, particularly with reference to Huber products. "While the analyses and charts included in these pages refer to Huber Clays and the photographs illustrate clay production at Huber plants, this book is nevertheless intended as an authoritative text on the broad subject of clay and its applications."

The brief review of the manufacture of paper in Chapter 10 leaves something to be desired. However, this can probably be overlooked because of its brevity, but there are some inaccuracies.

The testing methods and use of clays in paper as a filler and as a coating and in various rubbers are covered. It seems as though more information on the use of dispersing agents in clays for the paper industry could have been given.

A good section on compounding and evaluation of rubber goods is included. It is regrettable that it is not emphasized that the extensive comparative data presented are based presumably on laboratory results. While these data will indicate trends, they may not be representative of mill conditions and results, particularly in cases where dispersion may be difficult.

Plates and cuts require a coated paper, highly calendered; for good reproduction and this has been accomplished, but at some sacrifice on reading comfort from the glare of the pages. The text might have been set on a coated paper with mat finish, thus making for easier reading.

Valuable Reference Book

Directory of Biological Laboratories. 4th edition. Burns Compiling & Research Organization, 200 Railway Exchange Building, Chicago 4, Ill. (1949) 140 pp. $3.00.

THIS is an alphabetical listing (according to states) of laboratories of the United States concerned with biological or biochemical investigations. It includes research laboratories, consulting laboratories, and those related to manufacturing processes. Personnel and professional data are given. A special section is the Equipment Annual & Materials Index with a Buyers' Guide Section including list of manufacturers and distributors of apparatus, materials, supplies, equipment, and instruments. Included in the District of Columbia is a general reference to United States Government Departments. This book is valuable for professional reference and for the technical library. The only adverse criticism of the work is in the setup where a laboratory listing, such as the Argonne National Laboratory, comprising 30 lines of descriptive and explanatory matter is in one paragraph and in type difficult to read.

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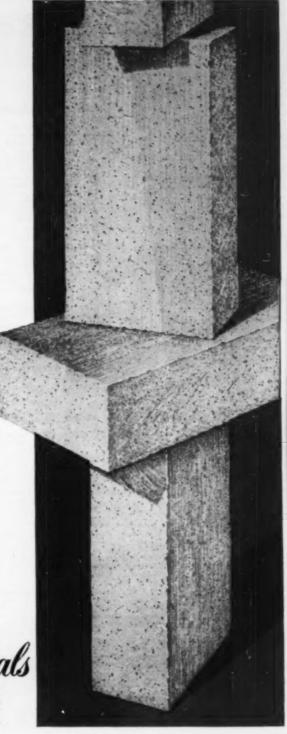
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SECRETARY'S PAGE

THE Executive Committee met Jan. 6 at the Offices of the Institute and after handling routine matters approved the election to membership of those whose names appear below. Several additional appointments were made to committees of the Institute.

The Council of the Institute met at the Hotel Commodore, New York, Jan. 6. Several matters were discussed but no final actions taken, therefore no report is made at this time. The question of the 1950 Year Book was discussed at some length and it was voted that a card be sent to all members with the Houston Meeting Program in an attempt to bring the Year Book information up to date. All members are urged to return these cards promptly.

Approval was voted for the establishment of the Columbia Valley Section of the Institute which would have its headquarters at Richland, Wash.

Request for approval of the Book Project had been received from CARE and it was voted that the Institute go on record as urging members to assist in this work to as great an extent as possible.

The Secretary reported on the organization of a Chemical Industry Correlating Committee by the American Standards Association and our representative on this Committee will be J. C. Lawrence and as alternate Wendell G. Fogg.

C.A.R.E. BOOK PROGRAM

The Institute has been requested to cooperate with UNESCO and CARE in providing technical books so seriously needed in the devasted and underdeveloped countries. The countries now served are Austria, Belgium, Czechoslovakia, Finland, France, Germany, Great Britain, the western zones of Germany, Korea, Japan, the Netherlands, Norway and the Philippines.

Cash donations are requested (do not send books).

Contributions of under $10.00 are pooled in a general fund. Donors of $10.00 or more can specify the book category, the country, specific institutions or type of institution which they wish helped.

Council has approved this activity, and it is hereby called to your attention. Contributions should be sent to CARE Book Program, 20 Broad Street, New York 5, N. Y.

THE Council of the Institute met Dec. 4 at the Hotel William Penn, Pittsburgh, Pa. The first item of business was to review the proposed amendments as they had been submitted to the membership for discussion at the Business meeting Dec. 5. An Employees Retirement Plan was approved as presented.

The Council approved the authorization of the Twin Cities Section of the Institute with headquarters at Minneapolis and St. Paul. The Chairman is Edgar L. Piret.

Upon recommendation of D. L. Katz, Chairman of the Student Chapters Committee, the Council voted to approve a student chapter at Lehigh University. Dr. D. E. Mack is the counselor of this student chapter.

Recommendations regarding committee personnel from the Chairmen appointed at the November meeting of Council were presented and personnel approved and it was directed that they be notified.

The budgets for both the Institute and *Chemical Engineering Progress* which had been presented at the November meeting of Council were further discussed and approved.

NEW A.I.Ch.E. MEMBERS ELECTED JANUARY 11, 1950

ACTIVE MEMBERS

A. J. Abbott
T. T. Anderson
Herbert A. Bahrenburg, Jr.
S. George Bankoff
Alexis G. Basilevsky
Norman H. Blumberg
Harry Blumenfeld
Stanford W. Briggs
Sherwin Chase
J. L. Christian
Richard I. Cope
Hal M. Hart
Walter S. Kaghan
W. A. Moggio
Robert W. Moorhouse
H. K. Nason
Robert E. Noble
Walter W. Northgraves
A. S. Parker
Edwin K. Plant
Vernon C. Quarles
Charles H. Riesz
R. C. Rohrdanz
Edward G. Scheibel
Martin D. Schlesinger
Willard F. Sheldon
C. Frederick Tears, Jr.
Walter G. Thomas
O. E. Toelle
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SECRETARY'S PAGE

The Council meeting was recessed subject to call of the chairman.

There was a reconvened meeting of Council Dec. 6 and certain modifications were made in the proposed amendments which were submitted to a subsequent Business meeting.

Upon recommendation of P. D. V. Manning, Chairman of the Professional Guidance Committee, it was voted that copies of "The Second Mile," a publication of the Engineers' Council for Professional Development, should be sent to all newly elected Junior members. This is a guidance pamphlet prepared by the late W. E. Wickenden and is of particular value to the recent graduate.

The Executive Committee met Dec. 13 in the headquarters office of the Institute. Bills for the month of November were approved and the Treasurer's report for that month received. New members elected as of Jan. 3 appeared on secretary's page, January issue, p. 30. Two elections to membership were rescinded because of nonacceptance. The following student chapter counselors were appointed as indicated:

M. L. Jackson at the University of Colorado to replace W. W. Wolf.

D. A. Dahlstrom at Northwestern University to replace G. Thodos.

There were a few additional committee personnel appointments which were confirmed.

The appointment of G. G. Brown as the Institute representative at the celebration of the Founding of the Wallberg Memorial Building of the University of Toronto was approved.

Resignations of 49 members were received and accepted.

Treasurer's Report

The financial position of the Institute as to cash and bonds as of Oct. 31, 1949, and compared with the same date the previous year is as follows:

	1949	1948
Total bank balances	$51,357.68	$33,325.50
Negotiable bonds, face value	25,000.00	25,000.00
U. S. Savings Bonds, at cost..	30,440.00	34,190.00
Total cash and bonds	$106,797.68	$92,515.50
Market value of negotiable bonds	$25,862.50	$25,168.76

On Feb. 1 of this year, $5,000 principal amount U. S. Savings Bonds, Series D, owned by the Institute matured. These cost us $3,750 and the difference of $1,250 represents accrued interest. There were no other changes during the year in the investments of the Institute.

C. R. DeLong, Treasurer

Only the new
HILLS-McCANNA
metering and proportioning pump

offers such complete dependability and freedom from maintenance

THE NEW Hills-McCanna "K" Type Pump offers a whole new conception of metering and proportioning. A simple straight-line enclosed hydraulic drive submerged in an oil bath replaces the complex mechanical linkages found in most pumps. Built-in overload relief makes it impossible to damage motor or mechanism by inadvertant overloading. Horizontal cylinders with twin checks reduce danger of air binding and can be easily disassembled for cleaning without dismantling piping. Stroke may be adjusted quickly and easily, from zero to full capacity, *even while the pump is operating.* Just compare and you'll see why "K" pumps are more dependable and require practically no maintenance.

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INST. GAS TECH.
OFFERS FELLOWSHIPS

The Institute of Gas Technology, as a part of its educational program, is offering 16 two-year fellowships effective September, 1950, to qualified college seniors and graduates. The purpose is to prepare a selected group for careers in the gas industry through graduate study, field training and research leading to the degree of Master of Gas Technology. Students in the upper fifth of their classes in chemistry, chemical or mechanical engineering or related fields may apply. Additional qualifications include United States citizenship, age under 28, adaptability, cooperation and high moral character. Fellows will be paid a cost-of-living stipend, currently $125 in each of the ten months of the academic year in addition to the remittance of tuition and fees amounting to approximately $550 per year.

The fellowships provide for the two years of study, including one summer training period and a thesis, which are required for the Master of Gas Technology degree.

Application forms and further information may be obtained from the student's departmental chairman or from the Director, Institute of Gas Technology, 3300 South Federal Street, Chicago 6, Ill.

I.U.T.A.M. ADMITS
U. S. COMMITTEE

The U. S. National Committee on Theoretical and Applied Mechanics, recently organized by seven national scientific and engineering societies, has been admitted by unanimous consent as an adhering body to the International Union of Theoretical and Applied Mechanics. This information was contained in a communication from Dr. J. M. Burgers, secretary of the Union with headquarters in Delft, Holland.

The U. S. Committee grew out of American participation in the Seventh International Congress on Theoretical and Applied Mechanics held in London, April, 1948. The need for some coordination of widespread activity in the field of mechanics in the U. S. prompted the applied mechanics division of the ASME to suggest a permanent organization of various national bodies to sponsor the U. S. for participation in international congresses and symposia.

Future plans call for American representation at a collegium on geophysics in 1950, and at a meeting of the IUTAM in Rome.

U. S. RUBBER BUYS
GLENN MARTIN CHEM. DIV.

United States Rubber Co. has purchased the chemicals division of The Glenn L. Martin Co., according to a joint announcement made late in December by the two companies. The rubber company will acquire all assets of the division, including the Marvinol vinyl resin plant at Painesville, Ohio, laboratory equipment in Baltimore, patents and the trade name Marvinol. The purchase price was not announced.

The business will become a part of the Naugatuck chemicals division of the rubber company. The Painesville plant will continue to make Marvinol vinyl resin for sale to manufacturers of plastic products.

Through the sale of its chemical division, the Martin Co. completes its program of concentration in the aviation field, by divesting itself of all interests not directly concerned with the manufacture of airframes, special weapons and closely related products.

United States Rubber Co. said the purchase is part of a long-range program to expand its activities in the chemical field.

The plant at Painesville, built in 1947, currently employs about 200 persons. It is close to raw material sources and has good transportation facilities.

U. S. TESTING, ESSELEN
RESEARCH MERGE

The merger of the United States Testing Co., Inc., Hoboken, and the Esselen Research Corp., Boston, was announced late in December, 1949. Under the plan the Esselen Research Corp. will become the Esselen research division of the United States Testing Co., Inc., and will continue its operations in Boston under the direction of Dr. Esselen, who founded the corporation. The merger will make available to industry a combination of research personnel, facilities, and experience.

Dr. Esselen is known as an authority on cellulose chemistry and his associates in the laboratory comprise a group with wide experience in many fields of industrial chemistry and chemical engineering. Their services include new product development, technical advice on legal matters and economic surveys.

The United States Testing Co., Inc., is one of the oldest and largest scientific testing organizations in the country. Among the fields it serves are the following: textiles, metals, ores, plastics, minerals, fuels, chemicals, etc. In addition it operates an engineering laboratory which offers a research and testing service in such fields as acoustics, X-ray, air-conditioning.

EQUIPMENT

1 • NEW HYDRAPULPER. The Cowles Co. has in production a new side drive pulper for the paper trade. Based on studies of pulping and defibering, the company designed a fast running side impeller circular pulper. The impeller is partially shrouded by a member resembling the volute of a centrifugal pump. This collects and directs the discharge from the impeller in one general direction. Impellers run from 200 to 500 rev./min. A 75-hp. pulper will pulp a ton batch in 10 minutes. Other engineering features of the pulper are fast dumping provided by a steeply sloping bottom, and a good submergence of the stock provided by an opening at the top of the shroud.

2 • LIQUID-LEVEL INDICATOR. The Asch Equipment Co. has introduced the Levlocator, an instrument for measuring the liquid-level content in a tank or storage vessel. Operates by means of an electrical or electronic finder, which operates a direct-reading indicator showing the amount of liquid above the preselected data point. According to the company it is accurate to 1/100 in.

3 • SYNTHESIZING UNIT. A packaged direct-fired synthesizing unit consisting of 1000-gal. agitated kettle, a 2000-gal. jacketed agitated cooling and thinning tank, plus automatic recording oil burners and a reflux condenser, is a new engineer unit of the Patterson Foundry & Machine Co. It is designed especially for oil-bodying, synthetic resin manufacturers, and other polymerizing, refining and chemical processing operations.

4 • NEW CENTRIFUGE. A new line of centrifuges giving external control of solids discharge and solids consistency, is announced by the Sharples Corp. The new line features 10 externally controlled solids discharge valves located radially

around the inside of the bowl. They are operated by water admission through separate channels. Can be used as a separator or clarifier; a bulletin is available.

5 • PRESSURE EXPANSION JOINTS. For high pressure industrial piping, Chicago Metal Hose Corp. has designed a new group of expansion joints for use up to 1000 lb./sq. in. Trade-marked Flexoniflex, the joints consist of corrugated bellows-type sections of stainless steel or other alloys, formed with integral control rings and end sections. Sizes begin at ⅝ in. I.D. up through the range of extra high pressure pipe sizes. Temperature range from subzero to 1400° F. for stainless steel, and units for higher pressures are available. Units for absorbing radial and offset motion, as well as axial motion, are company developments also.

6 • HEAVY DUTY VALVES. For pipelines and equipment handling paper pulp, thick or heavy chemicals, slurries, sludges, etc., the W. S. Rockwell Co. features new valves that may be cleaned of accumulated solids quickly and without removing them from the pipeline. Available for line pressures up to 100 lb./sq. in. The sediment is removed by swinging out of place a covered plate and dropping out the accumulation in the sediment chamber. High pressure air, steam or water connected

to the bonnet would clean out the entire valve. Cast iron and bronze bodies, plus steel and stainless steel blades, are in production and lever handle controls and rising stem controls are standard.

7 • FLOW METERS. Direct reading flow meters for measuring gas flow and available in several types of mountings, are announced by Hoke Inc. New "bantam"-type flow meters available for flow range 0.05 to 0.7 l. of oxygen/min., and another for 10 to 100 cc. of CO_2/min., are described by the manufacturer in a bulletin.

8 • MATERIALS HANDLER. A new, large capacity hydraulic machine for dumping liquids or solids into high level kettles, was recently developed by the Lee Metal Products Co., Inc. The handler is portable being mounted on rubber-tired casters, and barrels and bags may be mounted on the unit for lifting to the top of the tank. The drums or bags are placed in a stainless steel trough. A stainless steel rod is used to go inside a bag to keep it open and from slipping down when the trough is lifted and dumped. Trough is raised and lowered by means of a hydraulic cylinder operated from the city water line. Liquid mixes in cans or drums can be poured slowly into batch mixers. A bulletin describing the unit is available.

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Chemical Engineering Progress Data Service

I would like to obtain more information on the items represented by the numbers I have circled.

1	2	3	4	5	6	7	8	9	10	11	12	14	15	16
17	21	22	23	24	25	26	27	28	29	30	31	32	33	34
35	36	37	42	43	44	45	46	47	48	49	50	51	52	

Name .... Position ....

Company ....

Address ....

City .... Zone ...... State ....

9 • CRUSHER SIZER. A combination crusher sizer for operation on semi-friable material such as resins, plastics, clay, etc., is now on the market by Gifford-Wood Co. The unit will handle lumps up to 10 x 10 and crush and grade the material to three sizes. Capacity is 200 lb. of crushed material a minute. It is equipped with a bagging hopper and bag spreader to handle the final crushed product. Power is provided by a single 1½-hp. motor.

10 • MERCURY THERMAL SYSTEM. For control of temperatures above 1000° F. on gas-fired ovens, blanchers, etc., Taylor Instrument Co. has a new 1200° F. mercury thermal system. Thermal element is stainless steel. A new type Bourden spring is used on the indicator, and the new instrument is for use on systems beginning at 700° F. upward.

11 • STEAM BOILER. Eclipse Fuel Engineering Co. has a new gas-fired horizontal tube boiler for industrial plants. The shell of the boiler is a horizontal steel cylinder containing inside of it another steel cylinder which is the furnace in which the gas is burned. At the back of the furnace is an insulated chamber from which small, long return tubes conduct the furnace gases back the entire length of the boiler, so that the maximum amount of heat may be extracted. It is called the McKee Gas-fired Scotch Steam Pak. The

company describes the construction in a bulletin. It is built in nine different sizes ranging from 12 hp. to 100 hp. in accordance with boiler codes.

12 • PULSAFEEDER CONTROL. For automatic control of the Pulsafeeder proportioning pumps made by Lapp Insulator Co., a new pneumatic system has been devised by the manufacturers. The output from the Pulsafeeder pump can be adjusted while in operation and the new automatic control makes volume change from zero displacement to full displacement entirely automatic and in response to instrument air pressure change. The new control makes possible systems in which pump output is made responsive to any measurable variable such as flow, pH, temperature change, etc. Complete literature available.

14 • ATMOSPHERIC CONTAMINATION CONTROL. An infrared analyzer for the control of small quantities of contaminating gases in industrial discharges, is a new development of the Mine Safety Appliances Co. Trade-named Lira, the unit operates on the principle that most gases absorb infrared light and the mixture to be analyzed is introduced into a sample tube and its infrared absorption compared with that of a standard filter tube. Absorption is evidenced by pressure differential applied to a sensitive membrane and

the amplification of these movements by an electrical signal which registers on a voltmeter. The company claims that low gas concentrations may be detected within a range of 100 to 1 p.p.m. Suggested for measuring butane and propane in natural gas before and after processing; for measuring performances of petroleum industry units producing a specific hydrocarbon; for recording the content of process gases, water content, etc., in the industrial field.

15 • GAS MASK. A gas mask for use against hydrogen sulfide and petroleum vapors, meeting the approval of the U. S. Bureau of Mines, has recently been developed by the Willson Products, Inc. It features an oversized canister which provides protection for twice the period specified by the U. S. Bureau of Mines. Specially thought of interest to all refineries where sour crudes evolve hydrogen sulfide, processing and other industrial applications such as blast furnaces, cellulose manufacture, coke-oven workers, etc.

16 • UTILITY PUMP. A new utility pump for application in corrosive installations for the textile, chemical, paper, etc., industries, is newly designed by Burgess-Parr Co. Of a special corrosion-resistant alloy, the pump has a semi-open impeller driven at 3600 rev./min. Designed primarily for laboratory and pilot plant uses. Will deliver 30 gal./min. at a 40-ft. head.

17 • GAS FLOWMETER. A variable-gas flowmeter available in capacities ranging from 45 to 16,000 cu. ft. of CO_2/hr. (and comparable rates for other gases), is a new offering of the Fischer & Porter Co. The metering tube and float are fabricated from stainless steel. Extension below the metering tube carries a metering scale, and an indicator below the float permits direct reading of flow rate. Can be furnished to withstand pressures up to 600 lb./sq. in. gage. Operating temperature is limited by the properties of commercially available packings. Additional information available.

BULLETINS

21 • DIESEL DIGEST. A well-written and conceived 45-page illustrated book offered by the Sinclair Refining Co. gives an excellent insight into the history, economics, operation, etc., of the Diesel engine. In addition to the mechanical knowledge, the book contains a large section on fuel. One table gives the relation between fuel characteristics and engine performance.

22 • STEAM BOILERS. For those contemplating new steam systems, a new advertising brochure of the Cleaver-Brooks Co. will be informative. The units are "packaged," and are available in sizes from 15 to 500 horsepower and pressures from 15 to 200 lb./sq. in. Well illustrated. Shows, in diagram, the four-pass system of these boilers.

23 • ENTRAINMENT SEPARATOR. For removal of entrained liquid and solids from gases and steam, a complete line of Swartout separators has been designed and described in a 12-page bulletin. The separator uses a stationary helix near the inlet and the gases in passing are given a centrifugal motion. This whirl throws the unwanted heavier elements against the separator wall while the steam or gas continues through the line. Horizontal or vertical models in many models and pipe sizes.

24 • TYGON TUBING. An engineering catalog of the types, sizes, corrosion resistance, physical properties, etc., of a modified vinyl resin tubing manufactured by the U. S. Stoneware Co. The tubing is used in pilot plant, industrial operations, etc., where chemical resistance is important. Tables give the performance expectations of the tubing in various corrosive materials and the physical characteristics of the tubing after immersion in chemical test solutions. Uses are detailed and illustrated.

25 • CONTINUOUS EVAPORATOR. Bowen Engineering, Inc., describes a new evaporator for concentrating substances which have high fouling

characteristics. Originally developed by Sperti Foods and trade-named Stabilizer, the unit is a welded steel jacket, the inner wall being a horizontal cylinder in which a rotor turns. The rotor shaft is driven at low speed, the liquid to be evaporated passing between the rotor and the steam-heated jacket. Rotor motion causes a positive flow of liquid at a high surface velocity. Various materials of construction, for use where heat sensitive materials must be concentrated. Pharmaceuticals, foods, organic chemicals, etc.

26 • CORROSION-PROOF FANS. For moving gases and air conditioning corrosive materials The Duriron Co. shows in a published bulletin its line of corrosion resisting fans. Made of the various Duriron alloys (Durimet, Durichlor) the fans are made in sizes from 20 to 5000 cu. ft./min. Good dimension drawings of fans, plus capacity tables, method of mounting motors, etc. Applications are shown plus construction details, and alloy to use for particular corrosive condition.

27 • LEAK DETECTOR. An explanation of the mass spectrometer principle of leak detection published by the Consolidation Engineering Corp. Drawings show methods of using the equipment (probe and envelope) and then there is an extensive explanation of the performance, circuits and parts of the spectrometer.

28 • RUBBER-LINED FITTINGS. B. F. Goodrich issued an eight-page catalog section on rubber-lined pipe, fittings, valves, flanges, expansion joints, etc. Construction is shown in cut-away views.

29 • RUBBER-LINED TANKS. Also a Goodrich catalog, this shows the various rubber tank linings available. Tabulates the chemicals that may safely be used in tanks lined with a particular rubber. Has tank dimension table and brief mention of cost and economics of rubber lining.

30 • EVACTORS. A comprehensive catalog of Croll-Reynolds Evactors—steam jet ejectors for high vacuum—is newly printed. The catalog contains sections on the ejector vs the mechanical pump, methods of selecting the right size of ejector, descriptions of the various sizes, applications, layouts, etc.

31 • CHART OF RESISTANT CEMENTS. To aid the engineer who must select a cement for use in a corrosive environment, The Pennsylvania Salt Manufacturing Co. offers a chart showing the resistance to chemicals of both resin and silicate cements.

32 • PRESSURE VESSELS. A. O. Smith Corp. issues, with this bulletin, the first of six on the subject of multilayer pressure vessels. The bulletin covers the manufacture and assembly of the units, which are built by wrapping

FOR FREE DATA

ping, tightening, and welding successive layers of thin steel plate around a pressure tight cylinder. No limit to size of vessel. Illustrated.

33 • SAFETY WITH ClO₂. The National Board of Fire Underwriters has issued a research report on Precautionary Fire and Explosion Safeguards in the use of Chlorine Dioxide for Industrial Bleaching." The chemical is used in bleaching flour, fats, oils, sugar solutions, paper, etc.

34 • UNION CARBIDE AND CARBON. This huge corporation explains its companies and the products they make. Historical resumes and descriptions of the five major corporation groups—Alloys, Chemicals, Batteries, Industrial Gases and Carbide, and Plastics.

35 • ANTHRAFILT. The tenth edition of a manual on the use of anthracite coal as a filter material replacing sand and gravel. Palmer Filter and Equipment Corp., in this pamphlet, answers questions on the use of the material and gives data from installations in operation.

36 • TECHNICAL DATA. A Lefax catalog packet listing over 2000 titles of data sheets available, plus sample sheets and information on how to take notes, filing methods, indexing, etc. Covers mathematics, engineering, chemistry, etc.

37 • WASTE TREATMENT. For engineers requiring a complete catalog of equipment used in waste-treating, the American Well Works has published a 24-page manual. It covers activated sludge-screening, grit removal, pre-treatment, sedimentation, etc. In most cases details of the equipment are shown.

CHEMICALS

42 • 43 • POLYVINYL ACETATE. American Polymer Corp. has published two bulletins on polyvinyl acetate emulsions (42) and polyvinyl acetate solutions (43). (42) Has sections on theory of emulsions, properties, compounding, storage, etc. Uses of acetate emulsions are given plus graphs of various physical properties. (43) Contains data essentially the same as the above plus charts on viscosity, physical properties, evaporation rate, etc.

44 • CHOLINE CHLORIDE. For the feed industry Commercial Solvents Corp. describes the use of choline chloride as a livestock and poultry feed ingredient. Gives usual data on amount of supplement for feed, plus information on handling, mixing, etc.

45 • SOLVENTS. Technical descriptions of solvents manufactured by Pennsylvania Industrial Chemical Corp. are given in a 6-page brochure. Describes characteristics and physical properties of aromatic petroleum naphthas, coal-tar naphthas, solvent oils, etc. For use by paint and varnish, insecticide, organic chemical, leather, printing ink and other industries.

46 • FORMALDEHYDE. The Chemical Division of Celanese Corporation of America in a 30-page technical data book on formaldehyde has included all pertinent data necessary for engineers and production men handling this product. In addition to physical properties and methods of analysis, it covers uses, toxicology and first aid, storage and handling.

shipping and unloading. Illustrated, it contains many tables, and has pictures of correct method of coupling and uncoupling tank cars.

47 • VINYL BUTYRAL RESINS. For the plastic field Bakelite Corp. has issued Technical Release No. 11, a 22-page bulletin on vinyl butyral resins. Shows general properties, compatibility, solubility, applications, formulas, etc.

48 • GLYCERINE IN COSMETICS. A technical data book published by the Glycerine Producers' Association discusses use of glycerine in drugs and cosmetics. Shows how the four properties of glycerine as a humectant, solvent, sweetener, and antiseptic, fit into the requirements of these industries. Includes data and tables on specific gravity, viscosity, freezing point, relative humidity, etc.

49 • TRANSPARENT FILM ADHESIVE. For the users of transparent films, National Starch Products Inc. has a chart showing types of transparent tapes, how to identify the various kinds, who manufactures each kind, trade name, gages and thicknesses in which it is manufactured, and characteristics. There are 14 different films which are identified according to their basic chemical composition. The chart is part of a booklet on special adhesives made by this company for transparent films. The booklet tells how to use adhesives, applications, etc.

50 • BUTYL-META-CRESOL. A 14-page technical bulletin of the chemical division of Koppers Co., Inc., devoted to monotert-butyl-meta-cresol used as disinfecting agent. Bulletin contains representative formulas for germicides and disinfectants, plus complete information on the chemical itself. Typical formulas are given.

51 • SYNTHETIC DETERGENT. A nontechnical folder by the Antara products division of General Aniline & Film Corp. on synthetic detergents—their history and application. Designed for nontechnical users of detergents.

52 • CHOLINE DI-HYDROGEN CITRATE. For the pharmaceutical industry Chemo Puro Manufacturing Co. has a new form of the choline salt of di-hydrogen citrate. Especially adapted for tablet machines since the particle size is controlled.

LOCAL SECTION NEWS

NEW JERSEY

This section's first symposium and dinner meeting was held Jan. 17 at the Hotel Essex House in Newark, N. J. The afternoon technical meeting on the unit operation settling was attended by some 175 engineers, including a group of senior students from nearby Newark College of Engineering. Ralph Cohen of Hoffmann-LaRoche, chairman of the symposium committee, presided.

The problems involved in the clarification of liquids and the design of suitable equipment were discussed by Dr. E. J. Roberts of The Dorr Co. C. E. Lapple of the Du Pont Co. presented a review of the fundamentals of the motion of particles suspended in fluids. A new correlation for the prediction of pressure drop due to vapor flow through both fixed and moving beds of regular shaped materials was the subject of a paper offered by Dr. John Happel of New York University. The program was concluded with a paper by Dr. R. B. Booth of American Cyanamid, presented in his absence by J. S. Kennedy. The subject covered the history, theoretical aspects and application of froth flotation in the recovery of various mineral products.

Atomic energy was the subject of the dinner meeting, and Sumner Pike, member of the Atomic Energy Commission, drew the plaudits of 100 dinner guests with his discussion of the more recent developments and problems in the nuclear field, and their relation to the international situation. F. J. Van Antwerpen, editor, C.E.P., introduced Mr. Pike to the assemblage, which included Dr. W. L. McCabe, National President, and S. L. Tyler, Executive Secretary. Of particular interest to the group were Mr. Pike's comments on the pressing need for chemists and engineers to develop practical methods for cutting down waste in processing uranium and for recovering cheaply and efficiently that material already wasted.

Stephen L. Tyler, National Secretary, presented the official local section charter to this newly formed section at a meeting held in the Winfield Scott Hotel in Elizabeth, Dec. 13, 1949. Mr. Tyler expressed his pleasure at the rapid growth of this section, which, with its 375 members, is already one of the largest in the country.

Following Mr. Tyler's presentation, Charles J. Dodge, managing director, New Jersey Society of Professional Engineers, discussed the problem of the promotion of the professional status of engineers.

The technical program of the evening consisted of a discussion of the Houdriflow Catalytic Cracking Process by C. G. Kirkbride, vice-president and director, Houdry Process Corp. Significant among the more recent developments in the process is the use of the hydrocarbon charge vapors as the fluid medium in the lifting of the catalyst.

Reported by Donald A. Levenson

ST. LOUIS

This section held a meeting in the main ballroom of the York Hotel in St. Louis, Dec. 13, 1949, with 75 members and guests in attendance.

The address of the evening, given by Francis J. Curtis, then President of A.I.Ch.E., was entitled, "This Institute of Yours."

Right at the start Mr. Curtis told his listeners that members of a society get out of it benefits in exact proportion to what they put into it. Specifically on the subject of local sections, "the Institute's life," he believed that they could deal more with student chapters, chemical engineering education, professional training, personal advising and in following state legislation of interest to chemical engineers. He said an increase in membership activity was desired, and that those who worked hard on local committees are necessarily going to be advanced, as well as derive personal satisfaction. He gave statistics on all grades of membership and said "it is not the policy of the Institute to go out after quantity alone but it is the earnest desire that every properly qualified chemical engineer in the United States be a member, take part in its activities, and act as a creator of those activities."

Listing the activities of the Institute he gave first the publication and dissemination of chemical engineering knowledge. "We must build up a body of the best chemical engineering literature we can if the profession and science are to advance." He continued:

The second great activity of the Institute is in the personal contacts between its members. The prime method which the Institute uses to forward this end of its existence is the program of meetings. In 1946 there was inaugurated a program of regional meetings. The results have shown a big increase in attendance at national and seminational meetings.

Continuing on the Institute's activities Mr. Curtis said:

The Institute maintains high professional standards. The code of ethics is subscribed to by each member. The Walker and Junior Awards are given yearly for high excellence of chemical engineering work as reported in publications and recently there has been instituted the Professional Progress Award to recognize the broader professional success.

Mr. Curtis referred to the cooperation of the Institute with Engineers' Joint Council in the studies of the professional and economic status of chemists and chemical engineers; the accrediting system which set the standards and methods of examinations of schools. This system, he said, "has been adopted and extended to other branches of engineering by the ECPD."

Mr. Curtis then spoke of the work of the Committee on Student Chapters, the Committee on Admissions, Committee on Professional Guidance and other committees working on questions with regard to Constitution and By-Laws.

In summary he spoke as follows:

. . . the greatest field of enlargement seems to be in the activities of the local sections. Insofar as possible they should be a duplicate of the national society. We must personalize our local sections. We must see that people know each other, meet each other, and make friends.

Don't avoid criticism of the Institute but let it be constructive.

We feel that the Institute should have done a better job in explaining ourselves to the general public. The Public Relations Committee has taken upon itself this burden.

It was the consensus of the group that a periodic survey of Institute activities by some well-informed member of the Institute is highly desirable and necessary to its continued forward development.

Reported by R. S. Yates

ROCHESTER

At the Jan. 18 meeting 41 members and guests attended the dinner at Lorenzo's and 85 were present at the meeting at Rundel Memorial Library. The speaker was L. E. St. Johns whose topic was "Why Obtain a Professional License." The program was concluded by the showing of two movies: "Agitation" by the Du Pont Company and "Problem, Solution and Result," by Hercules Powder Co.

Reported by E. W. Costick

What every Chemical Engineer

should know about Pressure Filters

Pressure filters are probably the most commonly used filters in the processing industries and are, with few exceptions, intermittent or batch filters. Although there are many types of pressure filters on the market—disc, porous stone, cartridge, etc.—the plate-type and leaf-type are the most generally used for medium- and large-scale operations.

PLATE–TYPE FILTERS

The best known and most widely-used of the plate-type is the plate-and-frame filter press which usually employs filter cloth as a base for the filter cake. Also well known are the horizontal-plate filters which have certain definite advantages in small sizes for handling relatively small-batch operations, but which are high in cost and unnecessarily bulky when large filter areas are required.

LEAF FILTERS

There are many different types of pressure-leaf filters, such as the horizontal leaf, rotating leaf, and vertical leaf. Probably the simplest in design and most practical in application is the Niagara vertical leaf filter, employing all-metal filter leaves covered with permanent, fine-mesh wire filter cloth with openings small enough to quickly take a precoat of even the finest commercial filter-aids.

Limitations of this filter are that (1) It is a clarifying filter and cannot efficiently handle slurries with high percentages of solids; (2) It is not ideally suitable where the cake, not the liquid, is valuable.

However, it is ideally suited to most applications where plate-and-frame presses and other pressure filters are used for clarification purposes. Designed to combine most of the advantages of these other filters, it overcomes many of the disadvantages, particularly of the filter press, such as excessive time and labor for cleaning, poor cake-washing characteristics, bulky construction, uneven pressure distribution over total area, product leakage, non-uniform precoating, etc.

ADVANTAGES

Main advantages of Niagara Filters are:

(1) Totally enclosed, high-pressure filtration.
(2) Elimination of all labor and expenses connected with handling, washing, replacement, etc., of filter cloths.
(3) Higher rates of flow per unit filter area.
(4) Complete, sparkling filtrate clarity.
(5) Excellent cake-washing characteristics (almost true displacement washing).
(6) Greatly reduced floor space requirements.
(7) Rapid, easy cleaning and cake removal.
(8) 100% corrosion-resistant alloy metal construction . . . stainless steels, other metals if required, at reasonable cost.
(9) Easily jacketed for high-temperature operation.
(10) Maintenance almost eliminated, since there are no moving parts.
(11) One-man operation and handling, even of a battery.
(12) Unit filter area up to 500 square feet.
(13) Lower first cost where corrosion-resistant materials are required.

LOW OPERATING COST

This filter, although relatively new in its application to the chemical and processing industries, is rapidly replacing many of the traditional, old-style pressure filters. This trend is the result of an increasing tendency to use corrosion-resistant equipment and the realization by plant operators that lower equipment operating and maintenance costs must be achieved to keep profits up in the squeeze between increased labor and material costs and an increasingly competitive price situation.

For additional data and information on the Niagara pressure-leaf filter, the new self-cleaning "Auto-Sluice" filter, and Niagara's pilot filter rental service, write Niagara Filter Corporation, 3091 Main St., Buffalo 14, N. Y.

FUTURE MEETING AND SYMPOSIA OF A.I.Ch.E.

Chairman of the A.I.Ch.E. Program Committee

George E. Holbrook

Room 7406 Nemours Bldg.,
E. I. du Pont de Nemours & Co.,
Wilmington, Del.

MEETINGS

Regional — Houston, Tex., Feb. 26-March 1, 1950.

Technical Program Chairman: M. C. Hopkins, Pan American Refining Co., Texas City, Tex.

Co-Chairman: K. A. Kobe, University of Texas, Austin, Tex.

Regional — Swampscott, Mass., New Ocean House, May 28-31, 1950.

Technical Program Chairman: R. L. Geddes, Stone & Webster Eng Corp., Boston, Mass.

Co-Chairman: E. R. Gilliland, Massachusetts Institute of Technology, Cambridge, Mass.

Regional — Minneapolis, Minn., Sept. 10-13, 1950.

Technical Program Chairman: E. L. Piret, Minnesota Mining & Mfg. Co., Minneapolis, Minn.

Annual—Columbus, Ohio, Dec. 3-6, 1950.

Technical Program Chairman: John Clegg, Battelle Memorial Institute, Columbus, Ohio

Regional—Kansas City, Mo., May 13-16, 1951.

Technical Program Chairman: Walter W. Deschner, J. F. Pritchard Co., Kansas City, Mo.

Annual — Atlantic City, N. J., Chalfonte-Haddon Hall Hotel, Dec. 2-5, 1951.

SYMPOSIA

Indoor vs. Outdoor Plant Construction

Chairman: J. R. Minevitch, E. B. Badger & Sons Co., 75 Pitts St., Boston, Mass.

Meeting—Not scheduled.

Chemical Engineering Fundamentals

Chairman: Mott Souders, Shell Development Co., San Francisco.

Meeting—Minneapolis, Minn.

Chemical Engineering Kinetics

Meeting—Minneapolis, Minn.

Chlorine-Alkali Industry

Chairman: C. F. Prutton, Mathieson Chem. Corp., New York, N. Y.

Meeting—Houston, Tex.

Fluidized Systems in Chemical Syntheses

Chairman: E. R. Gilliland, Massachusetts Institute of Technology, Cambridge, Mass.

Meeting—Swampscott, Mass.

Ultrasonics

Chairman: Dudley Thompson, Virginia Polytech. Inst., Blacksburg, Va.

Meeting—Swampscott, Mass.

Chemical Engineering in Food Industries

Chairman: W. L. Faith, Corn Products Refining Co., Argo, Ill.

Meeting—Minneapolis.

Applied Thermodynamics

Chairman: James Coull, University of Pittsburgh, Pittsburgh, Pa.

Meeting—Not scheduled.

Phase Equilibria

Chairman: W. C. Edmister, Carnegie Inst. of Tech., Pittsburgh, Pa.

Meetings—Houston and Minneapolis.

Chemical Engineering in Glass Industry

Chairman: F. C. Flint, Hazel-Atlas Glass Co., Washington, Pa.

Meeting—Not scheduled.

Processing of Viscous Materials

Chairman: W. W. Kraft, The Lummus Co., New York, N. Y.

Meeting—Not scheduled.

Authors wishing to present papers at a scheduled meeting of the A.I.Ch.E. should first query the Chairman of the A.I.Ch.E. Program Committee, George E. Holbrook, with a carbon copy of the letter to the Technical Program Chairman of the meeting at which the author wishes to present the paper. Another carbon should go to the Editor, F. J. Van Antwerpen, 120 East 41st Street, New York 17, N. Y. If the paper is suitable for a symposium, a carbon of the letter should go to the Chairman of the Symposia, instead of the Chairman of the Technical Program, since symposia are not scheduled for any meeting until they are complete and approved by the national Program Committee. Before authors begin their manuscripts they should obtain from the meeting Chairman a copy of the Guide to Authors, and a copy of the Guide to Speakers. The first book covers the preparation of manuscripts, and the second covers the proper presentation of papers at A.I.Ch.E. meetings. Presentations of papers are judged at every meeting and an award is made to the speaker who delivers his paper in the best manner. Winners are announced in *Chemical Engineering Progress,* and a scroll is presented to the winning author at a meeting of his local section. Since five copies of the manuscript must be prepared, one should be sent to the Chairman of the symposium and one to the Technical Program Chairman of the meeting, or two to the Technical Program Chairman if no symposium is involved and the other three copies should be sent to the Editors' office. Manuscripts not received 70 days before a meeting cannot be considered.

LOCAL SECTION NEWS

PITTSBURGH

A comprehensive analysis of the management of technical operations was provided by Dr. Frank K. Schoenfeld, vice-president-technical, B. F. Goodrich Chemical Co., in a talk given at the Jan. 4 dinner meeting held at the College Club of Pittsburgh.

In discussing organization problems, Dr. Schoenfeld emphasized the importance of team effort, and the training, conditioning, and selection of suitable personnel. Selection of able group leaders who recognize the trusteeship and basic management responsibilities of their position in the organization was shown to be most important. The speaker defined corporate responsibility and explained how it affected the operations of technical management.

Reported by Hugh L. Kellner

PHILADELPHIA-WILMINGTON

A successful and interesting meeting was held Jan. 10 at the Clubhouse Hotel, Chester, Pa., at which time a sound-motion picture in color of "Visual Studies of Bubble Cap Tray Operation" was shown, through the courtesy of C. F. Braun & Co., Alhambra, Calif., by Clifford C. Burton and William S. Sevier, district engineer, New York office, C. F. Braun Co.

This film showed the range of liquid and vapor rates within which operation was considered satisfactory, and also the performance outside this range as affected by such conditions as coning, dumping, high liquid gradient, flooding and excessive entrainment.

Attendance was 120 at dinner and 210 at the meeting.

Reported by R. B. Chillas, Jr.

CLEVELAND

New officers who were elected at the November, 1949, meeting are:

Chairman—Dr. J. L. Wallace, The Glidden Co., 1396 Union Commerce Building, Cleveland

Vice-Chairman—Dr. T. J. Walsh, Case Institute of Technology, Cleveland

Treasurer—W. M. Gaylord, National Carbon Co., 11709 Madison Ave., Cleveland

Secretary—E. N. Marsh, The Standard Oil Co. (Ohio), Standard Building, Cleveland

Executive Committee Member—A. J. Teller, Fenn College, 1983 East 24th Street, Cleveland

Reported by E. N. Marsh

Piping equipment for every need
from one complete line...CRANE

Take This HCl Absorber Installation, for example. Here Crane equipment controls the flow of hydrochloric acid. But regardless of the fluids you may be handling, the valves, fittings, pipe and accessories you need are available from Crane. One order to your local, well-stocked Crane Branch or Wholesaler covers everything for the job . . . in brass, iron, steel and corrosion-resistant alloy materials.

To rely on the completeness of this **Single Source of Supply** is to simplify all piping procedures—from original design to final erection and maintenance. To put **Complete Responsibility** on Crane for materials avoids delays, assures better installations. And to count on the **High Quality** of every item from Crane is to help yourself to better piping performance—from end to end of every system.

CRANE CO., 836 S. Michigan Ave., Chicago 5, Ill. *Branches and Wholesalers Serving All Industrial Areas.*

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RESPONSIBILITY
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Process piping at Hydrochloric Acid Absorber unit. Crane supplies all piping equipment.

GLOBE VALVES

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DIAPHRAGM VALVES

GATE VALVES

BOLTS AND GASKETS

CHECK VALVES

PIPE

FLANGES

SCREWED FITTINGS

CORROSION-RESISTANT VALVES

FLANGED FITTINGS

FOR HOT OR COLD HYDROCHLORIC ACID *in concentrations up to 35%, and for many other corrosive services—Crane recommends No.1615 Neoprene-lined iron body diaphragm valves. Separate disc and diaphragm construction assures longer diaphragm life, positive shut-off of flow even should diaphragm fail. Y-pattern body design has greater flow-through capacity. Working pressure: up to 150 pounds water, oil, air, or gas; 180° F. maximum temp. Sizes: ½ to 6-in. Write for Folder AD-1761.*

EVERYTHING FROM ...
VALVES • FITTINGS
PIPE • PLUMBING
AND HEATING
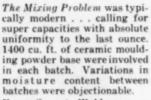
CRANE
FOR EVERY PIPING SYSTEM

UNIFORM BLENDS
By the Carload!

The Mixing Problem was typically modern . . . calling for super capacities with absolute uniformity to the last ounce. 1400 cu. ft. of ceramic moulding powder base were involved in each batch. Variations in moisture content between batches were objectionable.

From *Sprout - Waldron* came *the final solution*. First, six standard vertical mixers were each expanded to 300 cu. ft. These in turn were laced together with an ingenious system of screw conveyors, percentage feeders, and automatic controls. Plenty complicated — but it was all within the scope of Sprout - Waldron know-how.

Whether your blending operations involve uniformity by carloads or grams, consult Sprout - Waldron first. A wealth of practical engineering experience and broad manufacturing facilities are at your disposal.

● Bulletin 44 tells the Sprout - Waldron Equipment Story. Write for it today! Sprout, Waldron & Co., Inc., 17 Waldron Street, Muncy, Pa.

Sprout-Waldron
Manufacturing Engineers
MUNCY · PENNSYLVANIA

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LOCAL SECTION NEWS

AKRON

Approximately 60 members and guests attended the first meeting of the new year which was held Jan. 12, 1950, at the Elks Club.

R. P. Lowe, president, Proportioneers Inc., spoke on "Proportioning by Weight—By Volume." Mr. Lowe aptly described the fundamentals underlying fluid and solid metering. By means of slides he showed applications of automatic charging of liquids to Banbury mixers used in the manufacture of rubber products. A high degree of accuracy in proportioning of liquids by volume for continuous processes is achieved by a new development which utilizes an integrator having a true zero.

A regular meeting of this section will be held March 9. The speaker will be R. C. Gunness, manager, research, Standard Oil Company of Indiana, whose subject will be Synthetic Fuels.

Reported by John W. Kosko

OHIO VALLEY

Members of this section met in Cincinnati Jan. 5, 1950, to hear Edward A. Ulrich, vice-president and technical director, Niagara Filter Corp., discuss Filtration. Mr. Ulrich is an authority on filtration and clarification in the brewing and sugar industry. In the business meeting the section discussed the revisions to the local section by-laws.

Reported by A. C. Greber

WESTERN NEW YORK

This section held a joint meeting with the Niagara County Chapter of Professional Engineers Jan. 19 at the plant cafeteria of the electrometallurgical division of Union Carbide and Carbon Co., Niagara Falls, N. Y. The technical program consisted of a symposium on Dust Control and Elimination.

This section, a member of the Technical Societies Council, held a joint meeting with all technical societies in the Niagara Frontier Nov. 16, 1949. M. E. Streebe, president of the Long Lines Division of the Bell Telephone Co., described the mechanism of transmitting television, radio impulses from city to city. The growth of the coaxial cable mileage was traced to show how rapidly the industry is growing. Mr. Streebe demonstrated the microwave relay system of transmission on the stage. A film, "Stepping Along with Television," demonstrated how a program originating in New York was transmitted to Philadelphia, then Cleveland, Buffalo, Toledo and Chicago through coaxial cable.

Reported by E. C. Mirus

NORTHERN CALIFORNIA

At the Jan. 9 meeting 110 were in attendance to hear Paul H. Davies discuss the opportunities in American business and the relationships between government, management, and labor. President of Ford Machinery and Chemicals Corp., San Jose, Calif.

The newly appointed committee chairmen are:

Program—R. W. Lundeen
Arrangements—H. D. Guthrie
Publicity—J. V. Hightower
Professional, Economic Status—
                        J. A. Samaniego
Education—R. H. Ewell
Membership—A. George Stern
San Francisco Engineers Council—
                        N. P. Anderson
                        C. R. Nelson
            Reported by H. D. Guthrie

DETROIT JUNIOR GROUP

Officers of this group for the year 1950 were installed Jan. 9. They are:

President...........Leon Chrzan
Vice-President......Philip Lenton
Secretary-Treasurer
                Robert Cavanaugh
Program Chairman
                Richard Leslie
Membership Chairman
                Peter Bonnell

The speaker of the evening, Dr. Melvin Nord, chemical engineering department, Wayne University, addressed the group on the subject "Legal Aspects of Engineering." Dr. Nord pointed out the differences in the lawyer and the engineer and in the methods used by each. Law deals with people and concerns the application of social science whereas engineering concerns the application of natural science.

Dr. Nord illustrated his talk with interesting legal problems and in conclusion he pointed out several common legal entanglements with which the chemical engineer might easily become involved during the course of his career.

            Reported by P. A. Lenton

TWIN CITIES

The Minnesota Chemical Engineers Club was recently granted recognition as a local section and is now known as the Twin Cities Section of A.I.Ch.E.

The new section held a dinner meeting on Jan. 25, 1950, in Minneapolis. After the business meeting M. H. Baker, market research department, General Mills, Inc., gave a talk on "Chemical Process Development."

            Reported by W. M. Podas

NEW ORLEANS

This section, and the Delta section, A.I.M.E., jointly sponsored a technical session of the Louisiana Engineering Society 1950 annual meeting at New Orleans. At the session, Jan. 14, A. L. Vitter, Jr., petroleum engineer, the California Co., spoke on "Offshore Operations."

In 1948 the Louisiana Council of Engineers (this section is a member) succeeded in having a professional engineers' registration act passed by the legislature. This act was subsequently vetoed by the governor. On Jan. 6, 1950, the Louisiana Council of Engineers met again to consider further action in this matter. The meeting was well attended with representatives from the various engineering societies. It was brought out that those who requested a veto of the bill in 1948 had been approached and that misunderstandings had been clarified. By acclamation it was decided to make an effort to reintroduce the 1948 version of the bill into the 1950 legislature if an assurance could be obtained from the governor that it would not be vetoed again. Such an assurance appears likely.

At a meeting Jan. 19 J. H. Rushton, director, department of chemical engineering, Institute of Technology, Technology Center, Chicago, Ill., spoke on "The Technology of Mixing."

Reported by R. M. Persell, and K. M. Decossas

CHARLESTON, W. VA.

This section held a meeting Jan. 17, 1950, at the North Charleston Recreation Center. The speaker of the evening was Ragnar D. Naess of Naess, Thomas & Thielbar, investment counsel, of New York. His subject was "An Investment Manager Looks at the Chemical Industry."

Mr. Naess compared from an investment standpoint the stocks of nine chemical companies with the thirty stocks included in the Dow Jones Industrial Average. He expressed the opinion that the chemical stocks take first place among the few common stocks that are an absolute "must" for inclusion in investment portfolios. Features of the chemical industry emphasized by Mr. Naess in supporting his opinion were its preeminence in dynamic growth possibilities, its conservative price policy compared with many manufacturing industries whose profits have been lifted to abnormal levels by inflation of commodity prices, its conservative financial structure, and its favorable earnings and dividend record.

The next meeting of the section will be a dinner meeting to be held at the Daniel Boone Hotel, Feb. 21, 1950.

Reported by R. W. King

LETTER TO THE EDITOR

Humanitarian Service by What Yardstick?

Sir:

James B. Weaver on page 33 of the December *Chemical Engineering Progress* seems to be depressed about the amount of service he renders to his fellow-humans.

I believe that he is using two standards of measurement. He measures himself and his fellow-engineers by very strict standards against a mighty ideal. He measures doctors, lawyers, and ministers by much looser and more generous standards against something less than an ideal. His results overrate the doctors and lawyers and underrate the engineers.

The groups selected by Mr. Weaver have this in common—they attempt to get humans who are in trouble out of their troubles.

There is much emotion involved in spiritual matters and I believe ministers should be omitted from the discussion.

Doctors, i.e., doctors of medicine and dentistry, and lawyers are commonly called on after we are in some sort of trouble, and we are in the mood, at such times, to regard them as something extraordinary. This idea is fixed more firmly in our minds by the fact that a segment of the persons practicing medicine and law is extremely vocal in praise of themselves and their professions. If we engineers spread propaganda about ourselves and our profession with an equally free hand, it would require only a limited time to "educate" the public to our "virtues."

In matters of law and religion there is little we can do as engineers to aid our fellow-humans directly. In matters of health and better living, there seems to me to be a multitude of tasks that are awaiting attention of competent men. I might suggest as a starter for Mr. Weaver, the field of human nutrition and health, where problems involving anything from the soil itself through the cycle of plant and animal crops, food and food processing, to the utilization of sewage, are all calling for competent attention. There are other fields.

We, as a profession, have the work of bettering the lot of our fellow-humans. Generally we have to work as teams, rather than individuals.

Again, may I suggest that Mr. Weaver evaluate his and other professions by similar standards.

Dec. 28, 1949.

A. M. Cox
Pacific Grove, Calif.

PERSONALIA

J. T. COX, JR.

C. S. Williamson, Jr., formerly head, school of chemical engineering, Tulane University, New Orleans, La., has been made editor of the *Sugar Journal*, New Orleans.

E. D. Kane, formerly associated with the Ontario Paper Co. Ltd., Thorold (Ont.), Canada, as process improvement engineer, is now with the Bell Aircraft Corp. in the rocket testing group at Niagara Falls, N. Y.

Arthur C. Pope, vice-president in charge of manufacturing, Ansul Chemical Co., was recently elected one of the four new directors.

A. P. WEBER

A. P. Weber has been named technical director of International Engineering, Inc., Dayton, Ohio, engineers and fabricators of chemical process equipment and industrial fans and blowers. He was formerly director of process development and process engineering, engaged in the nuclear energy and advanced scientific fields, for The Kellex Corp., New York, a subsidiary of The M. W. Kellogg Co., and Pullman, Inc.

J. T. COX, JR., WITH McBRIDE CONSULTANTS

Dr. John T. Cox, Jr., resigned his position as deputy director, Office of Rubber Reserve, Reconstruction Finance Corp., effective Jan. 1, to begin practice of consulting chemical engineering in association with Russell S. McBride and Gordon W. McBride at 1518 K Street Northwest, Washington 5, D. C. For some time he will also retain a consulting and advisory relationship to the Office of Rubber Reserve.

As director of production, Dr. Cox has been in charge of the government's synthetic rubber manufacturing program. He has been responsible for supervising the conversion of these plants to the manufacture of "cold rubber." Also under his supervision the government has developed liquid latex from synthetic rubber.

A native of Louisiana, Dr. Cox received his chemical engineering education at Louisiana State University and Massachusetts Institute of Technology, and was awarded a Doctor of Engineering degree by Case Institute of Technology for his work in the American synthetic rubber industry. After a few years in the sugar and paper industries, he was with Firestone Tire and Rubber Co. to develop and manage its synthetic rubber work until he was engaged by R. F. C. for the wartime rubber program.

Dr. Cox, a member of A.I.Ch.E., has represented Reconstruction Finance Corp. on the President's Scientific Research Board. He is a founder and past president of Chemical Engineers Club of Washington, a member of A.C.S., and a board member of District Council of Engineering Societies.

Wendell G. Fogg, executive engineer, Air Reduction Co., is located in the New York office of the company at 60 East 42nd Street, New York 17, N. Y. The company recently moved its laboratory from Stamford, Conn., to Murray Hill, N. J.

Erwin Schwenk is now research associate of the Worcester Foundation for Experimental Biology, Shrewsbury, Mass. He is also a consultant with the Schering Corp., Bloomfield, N. J.

F. G. BREYER HONORED BY AM. INST. CHEMISTS

Frank G. Breyer, member of the New York consulting metallurgical and chemical engineering firm of Singmaster & Breyer, has received honorary membership in The American Institute of Chemists by vote of its Council. Presentation was made at a meeting of the Institute's New York Chapter held Feb. 15 at the Hotel Commodore. Mr. Breyer was honored for his important applications of chemistry and engineering in industry. His acceptance address was titled "Grandpa and the Looking Glass."

F. G. BREYER
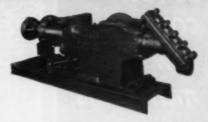
Mr. Breyer's professional career has been largely identified with the consulting firm of Singmaster & Breyer of which he was one of the founders. His outstanding contributions have been in nonferrous metallurgy, and in the related fields of application of these products. These include pigments, paints, inks, rubber and the graphic arts, in both their economic and processing phases.

Mr. Breyer studied chemistry at the Johns Hopkins University under Dr. Ira Remsen, receiving the degrees of A.B. (1908) and M.A. (1910). After graduation he went with the New Jersey Zinc Co., where he first met and collaborated with J. A. Singmaster. This association has continued uninterrupted ever since, and in 1927 the two men withdrew from New Jersey Zinc to form their own consulting engineering firm.

Arthur C. Ellsworth has been transferred from the Columbia chemical division of the Pittsburgh Plate Glass Co., at Barberton, Ohio, to the development department at the firm's Natrium (W. Va.) plant. Mr. Ellsworth has been associated with the firm's chemical division since graduation from the University of Denver in 1940. He holds the degree of B.S. in chemical engineering.

CLASSIFIED SECTION

Advertisements in the Classified Section of Chemical Engineering Progress are payable in advance, and are placed at 15c a word, with a minimum of four lines accepted. Box number counts as two words. Advertisements average about six words a line. Members of the American Institute of Chemical Engineers in good standing are allowed one six-line insertion free of charge per year. More than one insertion to members will be made at half rates. In using the Classified Section of Chemical Engineering Progress it is agreed by prospective employers and employees that all communications will be acknowledged, and the service is made available on that condition. Boxed advertisements one-inch deep are available at $15 an insertion. Size of type may be specified by advertiser. In answering advertisements all box numbers should be addressed care of Chemical Engineering Progress, Classified Section, 120 East 41st Street, New York 17, N. Y. Telephone ORegon 9-1560. Advertisements for this section should be in the editorial offices the 25th of the month preceding the issue in which it is to appear.

SITUATIONS OPEN

Chemical Engineer—Pilot plant and development work. Eastern location. 5-8 years' experience, preferably man with great initiative, resourcefulness and qualities of leadership. At least M.S. degree desired. Give details education, personal data and salary requirement. Box 14-2.

Chemical Engineer Supervisor—Attractive opening eastern location for pilot plant supervisor with broad experience organic chemicals. Please submit complete resume and salary desired. Box 15-2.

EXECUTIVE DIRECTOR OF TECHNICAL OPERATIONS
NEEDED

A multi-division corporation operating in the hydrocarbon and silica fields seeks an outstanding man to direct its production, engineering and research activities.

He will have full responsibility for plant design, engineering and construction, and the evaluation and control of both long and short-term research projects through laboratory, pilot plant, process and general engineering stages. In staff capacity he will also serve as senior executive on matters relating to

manufacturing, including process development and industrial engineering.

Our operations are varied and complex. Our record of growth and progress is outstanding. We are interested in further diversification along related lines with the result that our future growth potential is excellent.

Salary is commensurate with the responsibility of this top level executive position. Applications will be treated in strict confidence and will receive immediate consideration.

Box 18-2

SITUATIONS WANTED
A.I.Ch.E. Members

Unit Process Engineer—Sc.D., Ch.E., married, 27. Two years' successful industrial experience. Excellent academic record and references. Desire development and research position. Box 1-2.

Chemical Engineer—Ph.D., age 34. Thirteen years of diversified and responsible industrial experience covering catalytic cracking, complete refinery design and revamp, atomic energy design, light-ends recovery and oxygenated compounds separation. Strong in distillation, absorption, extraction and adsorption. Experience ranges from process design and development to project engineering. Desire responsible position requiring technical, administrative and organizational ability. Box 2-2.

Chemical Engineer—B.Ch.E. 1948, single. Experience in operation, construction and maintenance; vacuum distillation and drying, esterification, condensations, heavy emphasis on chlorination of benzene, phenol and ethyl alcohol. Box 3-2.

Chemical Engineer—B.S.Ch.E. 1943, 29, family. Three years in charge of automatic control instrument installations, maintenance. Four years' laboratory and pilot plant development of heavy metal soaps and esters with some plant layout

and design work. Available at short notice. Location immaterial. Box 4-2.

Chemical Engineer—B.Ch.E., 1940, married, 30. Five years' experience as process engineer in petroleum research and development and five years as analytical chemist in the insecticidal field. Desire production or development. Box 5-2.

Chemical Engineer—B.Ch.E., 1941, married, 31. Four years' experience in technical department, now chief chemist. Four years experience as troop commander in W. W. II. Desire production or technical department position. Box 6-2.

Chemical Engineer—With experience in petrochemical process development. 3½ years production in arsenicals and heavy chemicals. 2 years Naval engineering officer, destroyers. B.S.Ch.E. 1942, Tau Beta Pi, married, 29. Now in California. Box 7-2.

Sales Engineer—Ph.D. in Ch.Eng. Age 30, single, willing to travel. Professional engineering—New York. Five years' development and production experience in petroleum, fats and oils. Box 8-2.

(Continued from page 36)

Chemical Engineer — B.Ch.E., Rensselaer, 1938. Twelve years' industrial experience. Plant manager fish waste plant, supervisor penicillin process, supervisor nitric acid plant. Considerable start-up experience. Desire responsible production position. Write for resume. Box 9-2.

Chemical Engineer—31, M.S. Desire opportunity in production or process engineering West or Gulf Coast. 2½ years process studies and design with oil and chemical companies; 3 years Navy engineering officer; 1 year inspection supervisor; 1½ years glass production. Box 10-2.

Chemical Engineer — B.S.Ch.E., 1941. Six years' experience in design and construction of research equipment; two years project work. Desire location Pacific Northwest only. Box 11-2.

Sales Engineer — Aggressive, sales-minded. B.S.Ch.E. 1936, married, 38. Fourteen years' experience chemical and processing machinery industry, production, development, research, sales. Desire to represent you for Wisconsin area. Box 12-2.

Process Engineer—M.I.T., B.S., M.S., 1931, Chemical Engineering. Age 40, married, 18½ years with 3 major contractors, petroleum refinery and chemical plant equipment design. Desire responsible position with progressive company. Box 13-2.

Chemical Engineer—28, B.S.Ch.E. with distinction University of Minnesota. Six years' petroleum refining experience: fluid catalytic cracking, catalytic reforming, polymerization, distillation, and extractive distillation. Group leader in production technical service and process design. Experienced in evaluation and improvement of operations, report writing, data correlation. Desire change in location, preferably not South or Southwest. Box 16-2.

Versatile Chemical Engineer—30, B.S.Ch.E., Veteran, Tau Beta Pi. Three years of plant and equipment design and supervision of design of oil refineries and chemical plants. Three years of research and development and pilot plant work. Enthusiastic and cooperative worker. Box 17-2.

Reward—To the executive who needs for his staff a personable, ambitious chemical engineer of superior ability, I offer nine years of exceptional experience in process improvement, cost analysis, savings studies, design, and production in heavy inorganics, acids, insecticides, phosphates, phosphorus and its compounds. Box 19-2.

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Process Chemical Engineer—Specialist in high vacuum, development, operation, supervision. B.S. in Chem., 1939, Polytechnic Institute of Brooklyn, with graduate work in organic. Production supervision experience in TNT, Vitamin A and E. Excellent references. Box 21-2.

(Continued from page 35)

GILLENWATER RECEIVES W. VA. WATER POST

Lyhle E. Gillenwater has been appointed junior engineer for the West Virginia Water Commission according to a recent announcement.

Mr. Gillenwater holds an M.S. and B.S. in chemical engineering from West Virginia University. During his year of graduate work at the university, he studied a photoelectric method for measuring the dust content of coal. This study was made possible through a fellowship from the Calcium Chloride Association. His immediate concern will be with the commission's problem of coal washeries which cause pollution in the state's streams.

Mr. Gillenwater attained the rank of first lieutenant in the Army during World War II.

Robert F. Seubert is now senior technical representative with the Koppers Co., Inc., Pittsburgh, Pa. He is associated with the chemical divisions sales.

Paul Bame is now located at Antioch, Calif., with Fibreboard Products, Inc., as control chemist. He was formerly associated with the Valite Corp., Lockport, La.
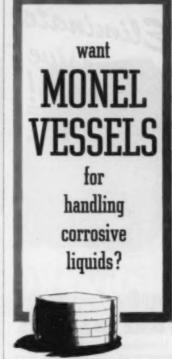
C. Leroy Carpenter is affiliated with the Colgate-Palmolive-Peet Co., Jersey City, N. J. He was formerly a graduate student in the department of chemical engineering, Polytechnic Institute of Brooklyn, Brooklyn, N. Y.

Robert B. Schow, formerly chemical engineer, research and development division, Diamond Alkali Co., Fairport, Ohio, is now production superintendent, National Distillers Chemical Corp.

R. B. Coleman, Jr., formerly associated with the Southern Research Institute, Birmingham, Ala., is now with the Southern Cement Co., in the same city.

R. Merton Shields has been appointed secretary of the Duriron Co., Inc., of Dayton, Ohio. Mr. Shields is a mechanical engineer, a graduate of Michigan College of Mining and Technology and prior to his new appointment was manager of the Dayton district sales for The Duriron Co. He has been with the company since 1936.

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PERSONALIA

E. A. ULRICH

Edward A. Ulrich was elected vice-president of the Niagara Filter Corp., effective Jan. 1, 1930. One of the original employees of the firm when it was founded in 1940, Mr. Ulrich has held various positions, and has been technical director since 1945. He will continue to hold that position in addition to his new office. With this company he has done considerable work in establishing markets for Niagara filters in the food, brewing, chemical, pharmaceutical and other process industries. A graduate of Purdue University in chemical engineering in 1931, Mr. Ulrich is the author of several published articles on filtration in the brewing and sugar industries.

J. Donald Charlton is now in Socony-Vacuum Oil Co.'s foreign trade department as process products engineer for Colombia, S. A. Mr. Charlton has been employed by Socony-Vacuum since 1942 except for two years in the Navy from which he returned in April, 1946, to the technical service department where he remained until June 1949.

TENN. EASTMAN MAKES TEXAS APPOINTMENTS

Managing personnel for the new Texas division of the Tennessee Eastman Corp., Kingsport, Tenn., to be located at Longview, Tex., was recently announced by James C. White, president of the corporation.

David C. Hull will be division manager. Mr. Hull has served Eastman in several executive posts since 1931. He was director of chemical production for Clinton Engineer Works, Tennessee Eastman Corp., Oak Ridge, Tenn., during World War II. He is a graduate in chemical engineering, University of Tennessee.

Other appointments include W. R. Saunders, superintendent of manufacturing operations.

The new Texas division will be located on a 2500-acre tract which was acquired by Tennessee Eastman, a subsidiary of Eastman Kodak Co. in the fall of 1949.

SWEARINGEN, ZINK, NEW DUTIES AT STANOLIND

Personnel changes in Stanolind Oil and Gas Co.'s Tulsa general office and other divisions became effective Jan. 1, 1950.

J. E. Swearingen, formerly executive assistant to the vice-president in charge of operations, has been appointed central division manager, and Dr. Donald G. Zink has been named general superintendent, chemical products department.

The new manager of Stanolind's central division, John Swearingen, joined Stanolind Oil and Gas Co. in April, 1947, as a project engineer, transferring from Standard Oil Co. (Ind.). In October, 1947, he became development superintendent of the manufacturing department and assistant to the manager of that department in 1948. He was made executive assistant to the vice-president in charge of operations, Jan. 1, 1949.

Dr. Zink was director of technical development for U. S. Industrial Chemicals, Inc., before going to Stanolind in 1947. As general superintendent of the chemical products department, Dr. Zink will be in charge of all commercial development and marketing in the field of chemicals.

Sidney Allan Savitt is now affiliated with the Consolidated Products Co., Inc., New York, N. Y., as a sales engineer. Dr. Savitt was formerly a consulting chemical engineer and prior to that was with The M. W. Kellogg Co., New York, N. Y., as a research engineer.

KIRKPATRICK GUEST SPEAKER AT ST. LOUIS

Sidney Dale Kirkpatrick, vice-president, McGraw-Hill Book Co., and editor *Chemical Engineering*, was guest speaker at a dinner at Hotel Sheraton, St. Louis, Jan. 20, climaxing dedication ceremonies of Washington University's new $500,000 Henry Edwin Sever Memorial Hall. The talk was titled, "Training for Leadership in Industry."

For 30 years Mr. Kirkpatrick has been closely connected with the development of chemical engineering. He has been associated with McGraw-Hill since 1921, first as an assistant editor of *Chemical & Metallurgical Engineering*, now known as *Chemical Engineering*, from 1921-24; associate editor from 1925-28 and has been editor in chief since.

Thomas B. Richey, Jr., formerly associated with the electrochemicals department, Du Pont Company, Niagara Falls, N. Y., is now associated with N. I. Malmstrom & Co., effective Dec. 15, 1949.

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